High Quality Effluents Seminar

Conference Proceedings No. 3

Research Program for the Abatement of Municipal Pollution under Provisions of the Canada-Ontario Agreement on Great Lakes Water Quality

PROCEEDINGS

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HIGH QUALITY EFFLUENTS

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FOREWORD

The Government of Canada and the Government of the Province of Ontario on August 13, 1971 entered into an agreement to ensure that the water quality of the Great Lakes is restored and protected. An important financial provision of the Agreement was the funding of a research program directed towards reducing the cost of waste treatment and ensuring that the latest technological advances were publicized to achieve the specific water quality objectives set out in the Agreement. A number of research projects have been funded since 1972, both in-house and by external contracts, to investigate and achieve high quality effluents from sewage disposal plants.

The Seminar on High Quality Effluents held in early December, 1975 was planned with two axioms in mind, namely: a particular environmental area can assimilate only a finite quantity of pollution; and, with the inevitable increase in the production of pollutants resulting from economic growth and urbanization, the degree of treatment must be improved.

In many instances, conventional treatment for the control of waterborne pollutants is no longer adequate and higher degrees of treatment are being sought. Lower levels of contaminating materials in sewage plant effluents are being required than can be achieved by secondary treatment.

The seminar was designed to discuss, in a practical and economic way, many of the questions concerning the need for higher quality effluents. The program was divided into three areas: environmental concerns; upgrading under conditions of hydraulic overload; and, upgrading under conditions of more stringent requirements. Topics discussed were surface water management, water resources assessment, wastewater treatment by disinfection, chemical addition, filtration and biological systems.

Results of research projects funded under the Canada-Ontario Agreement were discussed by approximately two dozen authors and researchers presenting 13 papers.

The attendance at the seminar was composed of over 200 registrants comprising approximately 20% from industry, 25% from the consulting sector, 15% from universities, 20% from the federal government and 20% from the provincial government.

The papers presented in these Proceedings should serve as a useful reference in the area of improving effluents from sewage disposal plants.

AVANT-PROPOS

Le 13 août 1971, les gouvernements du Canada et de l'Ontario concluaient un accord pour la restauration et le maintien de la qualité des eaux des Grands lacs. Une importante clause de cet accord concernait le financement de recherches visant à la réduction de coûts du traitement des eaux usées et stipulait que les découvertes techniques les plus récentes devraient être rendues publiques afin que les objectifs de qualité des eaux mentionnés dans le texte de l'accord fussent atteints. Depuis 1972, on a ainsi subventionné un certain nombre de travaux de recherches, entrepris par les gouvernements ou à contrat, pour l'obtention d'effluents d'usines de traitement des eaux usées qui soient d'un haut niveau de qualité.

Le Séminaire sur les effluents de qualité qui a eu lieu au début de décembre 1975 était fondé sur deux axiomes: un secteur de l'environnement donné ne peut assimiler qu'une quantité limitée de polluants; et il faux pousser le traitement, étant donné la hausse inévitable de la production de polluants, attribuable à la croissance économique et à l'urbanisation.

En maintes occasions, le traitement classique des polluants de l'eau ne suffit pas et l'on cherche à l'accentuer davantage. La pureté que l'on cherche à obtenir pour les effluents des installations de traitement est supérieure à celle que peut donner le traitement secondaire.

L'objet du Séminaire était d'aborder de façon pratique et économique les nombreuses questions que soulève la nécessité d'obtenir des effluents de meilleure qualité. Trois thèmes ont été touchés: la question de l'environnement; l'amélioration de la qualité dans les conditions de surcharge hydraulique; et l'amélioration de la qualité quand les exigences à cet effet sont plus strictes. Les discussions ont porté sur la désinfection, le traitement chimique, la filtration et les dispositifs biologiques.

Les résultats des recherches financées aux termes de l'Accord Canada-Ontario ont fait l'objet de 13 exposés présentés par deux douzaines d'auteurs et de chercheurs.

Parmi les 200 personnes inscrites au séminaire 25% venaient de bureaux-conseils, 15% d'universités et 20% respectivement de l'industrie, du gouvernement fédéral et de gouvernements des provinces.

Les exposés devraient servir de documentation à ceux qui oeuvrent dans le domaine de l'amélioration de la qualité des effluents des installations de traitement.

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SURFACE WATER MANAGEMENT IN ONTARIO

by

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Whenever water resources are scarce, choices have to be made to ration them among competing uses and users. In Ontario, water is made available for various users through a mixture of arrangements and procedures based on common law and provincial legislation.

There are two primary dimensions of surface water as a valuable resource. One is its physical quantity which is useful in both withdrawal uses, such as domestic consumption, industrial processes and irrigation, and in-place uses, such as power generation, navigation, recreation and aquatic life. The other dimension is its capacity to absorb and disperse waste, which is commonly referred to as the assimilative capacity of the watercourse.

RIPARIAN RIGHTS

The common law doctrine of riparian rights is confined to access and use. The riparian has no property right in the water resources himself, except that portion of it which he appropriates and uses. He is not required to use water, in order to preserve his riparian right. However, riparians are limited in the amount of water they may use. "Ordinary use" of water, including amounts that are reasonably necessary for domestic and livestock needs, may be made by a riparian without regard to the effect on downstream users.

The most important provision of riparian law pertaining to water quality relates to the protection from activities of others who interfere with their right to claim water. All riparians are entitled to the natural streamflow essentially undiminished and unaltered in character or quality.

PROVINCIAL LEGISLATION

The common law rules involving riparian rights are modified by the water-taking permit system set out in the Ontario Water Resources Act. This Act places the responsibility for supervising all the Province's waters in the Ministry of the Environment. The permit system leaves "ordinary" users in their common law position without requiring them to obtain a permit. But all users, with a few exceptions, for other purposes whose works were constructed or expanded after March 29, 1961 and whose withdrawal exceeds 10,000 gallons per day must obtain permits. Permits in themselves confer no private right to water use.

On a given watercourse, permits are granted on a first-come first-serve basis subject to several constraints and priorities. First, high priority is attached to protection of the supplies of ordinary users. Second, other extraordinary users who already hold permits are usually protected.

A major constraint in issuing permits relates to what is termed as "natural functions of streams". The objective is to avoid making allocations which would impair the ecological balance of the watercourse.

As permits must often be issued with imprecise data about stream-flows, terms and conditions are attached to permits to protect downstream users during periods of inadequate streamflow. For example, permits may require reductions in withdrawal rates during periods of shortage, and some permits contain a scale of withdrawal rates ranging from the maximum rate to zero. If shortages develop, each permittee is required to reduce his taking.

The main legislative instrument for regulating water quality in the Province is also the Ontario Water Resources Act. The basic legislative device is the approval required by the Ministry of the Environment for any disposal work. Approval of the work depends upon its design performance which in turn is judged through the Guidelines and Criteria for Water Quality Management in Ontario. This approval procedure, thus, links effluent treatment and equipment standards with environmental requirements.

The Ministry has a variety of strong general powers to prohibit any waste discharge whatsoever. For example, the Minister can apply for

an injunction to restrain the discharge of any material which in his opinion threatens water quality. It is a policy of the Ministry to reserve the use of this power for emergency situations only. The Act also prohibits any person or municipality from discharging any material which may impair water quality. Water is deemed to be impaired if the discharges cause injury to any living thing.

Superimposed on the receiving water criteria, the Ministry aims at the best practical treatment or control adequate to protect, and whenever possible, upgrade water quality. Municipalities are required to develop waste disposal systems where individual private systems prove to be inadequate and cause water quality impairment. Primary treatment is now the minimum requirement, with secondary treatment and certain specific measures required where the assimilative capacity of the receiving water is less adequate.

Additional provincial legislation are the Environmental Protection Act, which stresses the environmental and sociological aims of pollution control policy, and the Environmental Assessment Act. This Act received on July 14, 1975, third and final reading by the Provincial legislative assembly and represents the cornerstone of environmental planning and protection in Ontario.

Other provincial statutes, such as the Lakes and Rivers Improvement Act and the Public Health Act, contain provisions that restrict the depositing of substances in water, but these are aimed at specific purposes. The Lakes and Rivers Improvement Act also regulates the construction and operation of dams and other flow regulation and diversion works. This Act is administered by the Ministry of Natural Resources.

A potentially important role in water use can be exercised by the regional conservation authorities which, to the present, have been concerned primarily with water projects and particularly flood control works involving dams and river channelization. More recently, their involvement in flood land management and land acquisition, with recreation being an important concern, has increased the importance of the role conservation authorities can play in water management. The authorities are organized under the Conservation Authorities Act which is administered through the Ministry of Natural Resources.

A feature of Ontario's administration of municipal and regional water supplies and pollution control is the heavy involvement of the Province in municipal water and sewage systems. Provincial policy provides for involvement in the financing, building and operation of local works. In order to encourage municipalities - particularly small ones for which the cost of service per household is usually higher - to construct water and sewage systems, the government has introduced financial assistance schemes. The ultimate objective of these schemes is to provide required water and sewage facilities for all.

FEDERAL LEGISLATION

With regard to federal involvement, there has been an absence of significant federal participation in Ontario water resource management apart from water boundary issues. The surge in public interest in pollution and environmental matters, however, has led the federal government to take a much greater interest in federal water resource management through the Department of the Environment and the Canada Water Act.

The Canada Water Act empowers the federal government to enter into agreements with provincial governments with respect to any waters in which there is a significant national interest in their management.

In addition, there is federal jurisdiction over certain uses of water, the most important being navigation and fisheries. The federal Parliament preserves the public right to use inland waters for navigation through the Navigable Water Protection Act which requires official permission for any obstructions to navigable water. The Fisheries Act provides the federal Minister of the Environment with powers to control the obstruction of fish streams. However, the Ontario Ministry of Natural Resources effectively manages the commercial and sports fisheries in the Province. Also, through the Fisheries Act, regulations are made prescribing substances and quantities or concentrations that are deleterious. The federal government has begun to pass discharge regulations, such as those for the pulp and paper industries, that make the most important federal contribution to water quality control. Generally, in Ontario, the provincial discharge requirements are more stringent than the federal ones.

With regard to the international boundary waters of the Great Lakes, the International Boundary Waters Treaty Act was enacted to ratify the International Waters Treaty of 1909 with the United States. The Act is aimed at preventing and providing the means for adjusting and settling disputes between Canada and the United States over the use of waters along their common border. Towards this end, the International Joint Commission was established. The main driving force for the commission as far as water quality control is concerned is the following statement contained in the treaty of 1909:

> "It is further agreed that the waters herein defined as boundary waters and waters flowing across the boundary shall not be polluted on either side to the injury of health or property on the other."

In 1972, Canada and the United States signed the Agreement on Great Lakes Water Quality, by which both countries agreed to implement specific programs to reduce pollution from municipal, industrial, agricultural, shipping and other sources. The water quality objectives recommended by the International Joint Commission were adopted in the Agreement. The Canada-Ontario Agreement signed in 1971 was a major tool for implementing Canadian commitment towards the improvement of water quality in the lower Great Lakes. Part of this Agreement is the program of research into pollution abatement problems, the results of which at least in part will be highlighted later on in the program of this Seminar.

A particular problem identified in the lower Great Lakes system was the eutrophication of waters. Regulations under the Canada Water Act have already been used to reduce the content of phosphorus in household detergents. In addition, a major thrust of the funds provided for in the Canada-Ontario Agreement was towards the control of phosphorus in municipal sewage effluents discharging in the lower Great Lakes basin.

The above general outline of common law and provincial and federal legislation which directly affect water use and water management is followed by a more specific detailed description of how the "Guidelines and Criteria for Water Quality Management in Ontario" are applied by the Ministry of the Environment.

POLICY GUIDELINES FOR WATER QUALITY MANAGEMENT IN ONTARIO

In 1967, the former Ontario Water Resources Commission (OWRC) announced its "Guidelines for Water Quality Management in Ontario". and published them together with water quality criteria for various uses. Application of these criteria to water used within drainage basins and the management of water quality to allow as many uses as practicable became the basis for defining pollution control measures.

The policy Guidelines reflect the fact that any watershed is characterized to a significant degree by its water uses and abuses to water quality. To have a multiplicity of uses and maintain water quality is a direct function of urbanization, "people pressure": i.e. people pressure for more housing, recreation, boating, swimming, fishing, hunting and aesthetics; pressure for water supply both public and industrial; pressure for waste disposal, both municipal and industrial; pressure for navigation; and, pressure for total and supplemental irrigation. As a result, the issue of single-purpose, as opposed to multi-purpose use, has essentially become academic with respect to this Province's water resources. The real question is how to manage the multiple use watershed areas, including land and water uses, while allowing as many beneficial uses as practicable.

The Policy Guidelines of 1967 formalized in Ontario the concept of management on a drainage basin or watershed basis. In many cases, the dominant factor in determining how to manage watershed areas is water quality. Each use, whether for municipal, industrial, agricultural or recreational purposes of fish and wildlife habitats can impose some pollutant burden or demand on the water quality level. For the development of a watershed management plant, the various existing or potential uses of water in a drainage basin are determined and water quality criteria selected for these uses. Objectives and/or standards of receiving water quality are then established to protect these uses and effluent requirements or use restrictions are set to achieve or maintain the desirable water quality.

Although water quality is in many cases a dominant water management problem, flooding, and streamflow regulation to allow for water supply, irrigation, the dilution of wastes, and protection against flooding and erosion, fishery management, etc. are all part of the complex interrelated problems which potentially require resolution. It is through the ever

widening concept of water management that drainage basin studies are not only interdisciplinary, but involve the many agencies which have legislative control of the various aspects and interests outlined earlier and involve the private, local sectors of the public through public consultation programs.

The development of a management policy for a water resource, taking into consideration the total river basin, is a formidable and expensive task. It not only involves a systematic analysis of the Total basin's water resources, but also includes a review of land uses and their impact on water quality or their requirements for water quality.

The complexity of water management on a basin-wide basis is well illustrated by the overall objective of the Thames River basin study which reads:

> "To develop guidelines for water management planning in the Thames River Basin which would ensure that an adequate quantity of water at a satisfactory quality is provided for the recognized water uses in the river basin at the lowest cost and that flood and erosion protection is provided consistent with appropriate benefit-cost criteria."

A particularly strong feature of provincial policy in Ontario is the central importance given to the quality of receiving waters, thus acknowledging the need for flexibility in determining allowable waste inputs in the light of the varying assimilative capacities of lakes and rivers. Of course, the assimilative capacity varies directly with water uses and their quality and quantity requirements. One criticism that can be levied against this approach of determining the assimilative capacity is that the use of technical criteria fails to take cognizance of the relative value of the various public and private uses.

My following talk dealing with the Thames River will provide an insight into the problem areas and procedures used by the staff of the Ministry of the Environment and the many other agencies involved, such as Housing, Treasury, Economics and Intergovernmental Affairs, Agriculture and Food, Natural Resources, Health, etc., the local Conservation Authorities and the public in general, in the establishment of simple or complex water quality management plans and, specifically, pollution control requirements for municipalities.

WATER RESOURCE ASSESSMENT TECHNIQUES

by

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One of Ontario's major challenges is the management to satisfy various use pressures of its relatively abundant supply of fresh water. Since the early days of urbanization and industrialization, Ontarians have taken it for granted that they can settle almost anywhere, build, expand and dispose their sewage and industrial wastes into the nearest watercourse and be assured that they will have a cheap supply of good, clean water for drinking and not have to travel too far to swim or fish in a crystal clear lake or stream.

This attitude is changing; we have begun to realize that many of our important rivers, lakes and groundwater aquifers have been, or are in danger of, being contaminated to the point where they become unacceptable for most of our desired uses.

The various levels of government, industry, the academic community and concerned individuals have recognized the problem and over the past few decades have taken great strides to develop and implement equipment and techniques to reduce the discharge of polluting materials and undertake remedial actions to begin to clean-up our watercourses.

This paper documents the evolution and current status of aquatic environmental survey and assessment techniques of the Ontario Ministry of the Environment. For the most part, attention is centred on assessment techniques employed for evaluating the surface water resources of river basins.

HISTORY OF WATER RESOURCE ASSESSMENT IN ONTARIO

Protection of the province's water resources rested in the hands of the Ontario Department of Health until the mid-1950's, the Ontario Water Resources Commission until the early 1970's and today, the Ministry of the Environment.

In the early days of environmental studies in the 1940's and 50's, investigators concentrated on bacterial, solid and organic (BOD) material discharges, usually from municipal or private sources and, in most

cases, limited evaluations of the "polluting" impact to public health and basic aesthetic aspects in the immediate vicinity of the discharge. General water quality surveillance programs were for the most part limited to one or two grab samples per year at convenient access points along the major river systems and at a few key locations along the Great Lakes shorelines.

Municipal pollution surveys were undertaken as problems were identified throughout Ontario to assess the impact of malfunctioning septic tank systems, industrial discharges and other sources of pollution. Such studies often led to the recommendation of a communal sewage collection and treatment system.

Waste treatment requirements as they related to the receiving capabilities of a watercourse were usually determined by calculating the streamflow to sewage dilution ratio. As a rule-of-thumb, secondary waste treatment and a streamflow to sewage ratio of five or six to one or an instream BOD_5 of 4 mg/l were considered acceptable and the minimum dissolved oxygen objective of the day, 4 mg/l, would likely not be violated.

A general strategy employed by the province was to construct secondary treatment facilities (activated sludge, trickling filters, waste stabilization lagoons, etc.) for municipalities discharging to inland watercourses and primary sewage treatment plants in communities along the Great Lakes.

While the survey and assessment procedures of the past, in comparison to today's technology, may seem to be crude and the basis for selecting waste treatment modes scientifically and economically unsound, the fact is that Ontario was among the leaders in this relatively new field and the engineers and sceintists who led the way placed this province well ahead of most other North American jurisdictions in providing sewage treatment facilities and abating pollution.

In the early 1960's the Province of Ontario through the Ontario Water Resources Commission recognized the need for establishing groups of water resource specialists who were specifically responsible for investigating the physical, chemical and biological impacts of wastewater discharges on water quality and quantity. While working closely with the

sanitary or industrial waste engineers who were primarily interested in the "in plant" treatment processes, these specialists initiated their investigations at the "end of the pipe", gearing waste treatment requirements to the level of water quality required to support various water uses.

Principal responsibilities of this new group of scientific and engineering staff included three major areas. Firstly, pollution problems were identified by locating the source, magnitude and impact on water uses of existing (and potential) wastewater inputs. Secondly, remedial requirements were defined by establishing guidelines setting forth acceptable waste discharge rates for the various categories of "polluting" materials and recommending waste discharge procedures to ensure adequate supply of good quality water that will satisfy the requirements for all desired uses.

Finally, staff were commissioned with responsibility of routinely monitoring the province's lakes, rivers and groundwater at selected locations to ensure that established waste loading guidelines, operating procedures and abatement programs were being carried out and that the desired water resource conditions were being achieved.

WATER RESOURCE ASSESSMENT ACTIVITIES

1. Surveillance and Monitoring

One of the first activities of the Water Resources Assessment Group was to establish a water monitoring network where samples would be collected on a routine monthly (or more frequent) basis and analysed for a wide variety of water quality parameters. Concurrent streamflow measurements are also taken at many stations. Sampling stations are established downstream from most of the significant sources of pollution to monitor impact on water quality. Stations are also located near the mouths of all rivers discharging to the Great Lakes so that the impact of river drainage on the Great Lakes can be determined. Other key locations such as recreational areas of the province are sampled to determine the current status and trends in water quality. In 1964, when the river monitoring program was initiated, about 180 stations were in the network. Today we have water quality information for 930 stations.

The collection of water samples for the monitoring program is essentially a manual operation. Staff working out of the Regional offices, assisted by staff of the various conservation authorities and samplers hired locally in remote areas of the province, collect grab samples at specific locations at scheduled times. In the field, the sampler measures and records dissolved oxygen and temperature and, at some stations, water surface elevations which are later converted to streamflow. Weather conditions and visual observations such as ice cover, oil slicks, turbidity, unusual algae growth, etc., are also recorded. Samples for laboratory chemical, physical, biological and bacteriological analysis are collected in appropriate containers, "fixed" or preserved as necessary, and shipped or delivered to Ministry of the Environment laboratories.

Results from the field and laboratories are collated in the Ministry's head office and stored in computer files. Computer print-outs are forwarded to the appropriate Regional office where the information is evaluated and made available to the public.

Robot monitoring stations have been used at a few locations. These units continuously record dissolved oxygen and temperature, and some units also record pH, conductivity and turbidity. The data generated are useful in identifying short-term fluctuations in water quality caused by waste discharge variation, photosynthesis and respiration, etc. The data are also extremely useful in water quality model calibration and verification. Robot monitoring stations, however, do have some drawbacks in that they are costly to purchase and maintain and somewhat limited in their capability to monitor a wide range of significant pollution parameters.

Based on the Ministry's experiences to date, it appears that the routine monitoring program will continue to operate manually and robot monitoring stations will be installed and maintained at a few locations where continuous data are required for waste discharge surveillance or mathematical model verification.

2. Site-Specific Assessments Using Steady-State Mathematical Models

At the same time as the general surveillance network was being set up in 1964, the O.W.R.C. began to employ basic mathematical modelling techniques to determine the natural self-purification capacities of receiving watercourses.

These modelling studies, each requiring a great deal of fieldwork, are undertaken in the affected reaches downstream from single or a combination of waste sources. Survey reaches can extend from 4 to 6 miles in the case of small sewage treatment plants on small streams, to 80 miles for some large industrial sources such as pulp and paper mills.

Field crews ranging from 4 to 15 people survey the river in detail. Physical information such as velocity, lateral and longitudinal streambed profiles, time of travel, streambed composition, river and riparian uses, etc. is collected for later use. Biological studies to determine effects on aquatic life and to provide a longer term indication of water quality are often included. The key study designed to collect the water quality information that, along with the physical data will form the basis of the model, takes the form of a 48 to 96-hour "around-the-clock" survey where samples are collected at stations throughout the survey reaches at three to six hour intervals. These intensive studies are usually carried out during the critical period in the summer when streamflows are lowest and water temperatures highest.

With the completion of the field investigations, the data collected are analysed and factors such as the rate of BOD satisfaction and the rate of reaeration are established.

These coefficients are then employed in a modified form of the Streeter-Phelps dissolved oxygen balance equation which in its simplest terms, reads:

$$\frac{dD}{dt} = K_1 L - K_2 D$$

where: D is the dissolved oxygen deficit

t is time of travel downstream

K_1 is the coefficient of BOD satisfaction

L is the initial oxygen demanding loading

K_2 is the coefficient of reaeration

The equation and modelling process is essentially nothing more than an algebraic sum of the sources and losses of dissolved oxygen in the river system downstream from a pollution source.

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In the first few attempts at mathematical modelling, considerations of sources of oxygen were limited to the level measured upstream of the pollution source plus atmospheric reaeration. The only sink of oxygen considered was BOD satisfaction.

As experience was gained in mathematical modelling and the process was accepted as a management tool, other terms were added to the equation such as sludge demand, photosynthesis, respiration and nitrification. The steady-state dissolved oxygen model now employed by the Ministry of the Environment is as follows:

$$\frac{\delta d}{\delta t} + \frac{V \delta D}{\delta t} = -KaD + KdL(x) + KnN(x) + S - P(t) + R$$

where:

D	=	oxygen deficit, mg/l
V	=	velocity of streamflow, ft/sec
t	=	time, days
x	=	distance, ft
Ka	=	aeration coefficient, day^{-1}
Kd	=	deoxygenation coefficient, day^{-1}
L(x)	=	carbonaceous oxygen demand as a function of x, given by $L(x) = Lo_e^{-Kr(x/v)}$
Lo	=	initial concentration of carbonaceous oxygen demand, mg/l
Kr	=	oxygen demand removal coefficient, day^{1}
N(x)	=	nitrogenous oxygen demand as a function of x, given by $N(x) = No_e^{-Kn(x/v)}$
No	=	initial nitrogenous oxygen demand, mg/l
Kn	=	nitrogenous oxidation coefficient, day^{-1}
S	=	benthic bacterial respiration, mg/l/day
P(t)	=	photosynthetic oxygen source as a function of time, of the form $P(t) = PmSin[\pi/p)(t-ts)]$ for daylight hours. A Fourier series expansion of P(t) is used.
Pm	=	maximum rate of photosynthesis production, mg/l
P	=	period of sunlight, days (fraction)
ts	=	time of sunrise, days (fraction)
R	=	algal respiration, mg/l/day

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It is quite evident that undertaking a water resource modelling study to develop and implement waste loading or river management guidelines can be extremely expensive and time-consuming. In cases where small sewage treatment or industrial discharges or other minor activities affecting water resources exist or are planned, full scale field and modelling studies are not always warranted. The Ministry has developed assessment procedures employing quality and quantity monitoring records, estimated reaction coefficients and experience gained from the larger studies. These assessments generally do not require field work and can be completed in a few hours or at most, a few weeks. To date, these simplified equations have proven to be quite adequate for such applications as determining discharge periods and rates for seasonal lagoons, the impact of small additions to sewage treatment plants, flow regulation requirements of small reservoirs, etc.

As with the constant full scale field survey and modelling technique development and refinement, the simplified assessment procedures are continually being reviewed and improved.

DEVELOPMENT OF SOPHISTICATED ASSESSMENT AND MODELLING PROCEDURES

As water resource assessment technology evolved it became quite clear, particularly in southern Ontario river systems, that the impact of individual sources of polluting materials could not be looked upon in isolation. As well as the easily identifiable wastewater loadings from sewage treatment plants or industry, other sources such as land runoff from urban and agricultural sources, erosion, construction activities and other man-made and natural sources were influencing water quality, quantity and use.

The 1967 policy guidelines for water quality control in Ontario clearly established that the water resources of the province must be preserved and where necessary restored to a level for all desired uses.

"....the use of water for assimilation and dilution of treated waste effluents must take into consideration the variety of uses including public, agricultural and industrial supply, recreation, aesthetic enjoyment and the propogation of fish and wildlife."

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The increasing complexity and recognition of new waste inputs affecting water quality and the Ministry of the Environment's mandate to preserve water for all uses made it essential that assessment and modelling procedures had to be expanded in scope, flexibility and complexity, so that waste loading guidelines, streamflow management procedures, instream structures or devices or combinations of these measures could be defined in much more precise terms than was needed previously.

1. Dynamic Simulation Modelling

Simulation is a relatively new technique for water resource evaluation and planning. It has recently been employed for the first time by the Ministry in the Thames River Basin Study and its application in regard to water resource management guidelines in the Thames system will be discussed in a subsequent presentation.

Simulation has an advantage over steady-state modelling in that it can take into account daily, weekly or seasonal variations in any of the physical, chemical or biological factors affecting water quality or quantity. A dynamic simulation model is composed of a number of inter-related models designed to provide predictive capabilities for any selected water quality parameter or set of parameters.

In the Thames River Study for example, the primary water quality output was dissolved oxygen. Therefore, a variation of the dissolved oxygen steady-state model discussed earlier became the principle modelling component supported by a number of sub-models designed to calculate short and long term variability of the components of the dissolved oxygen model. This permitted the calculation of resultant water quality under any given set of conditions or on a continuous basis throughout a season or year.

In a simplified graphical example, the water quality (i.e. dissolved oxygen) prediction component of a dynamic simulation model is presented in Figure 1. This figure indicates the general capabilities of this type of model. All the components shown are not necessarily employed in every study.

Each of the sub-models supporting the dissolved oxygen equation is based on data available through historical plant operations records,

FIGURE 1 — SIMPLIFIED SCHEMATIC OF INPUTS TO THE DISSOLVED OXYGEN PREDICTION PHASE OF DYNAMIC SIMULATION MODELLING
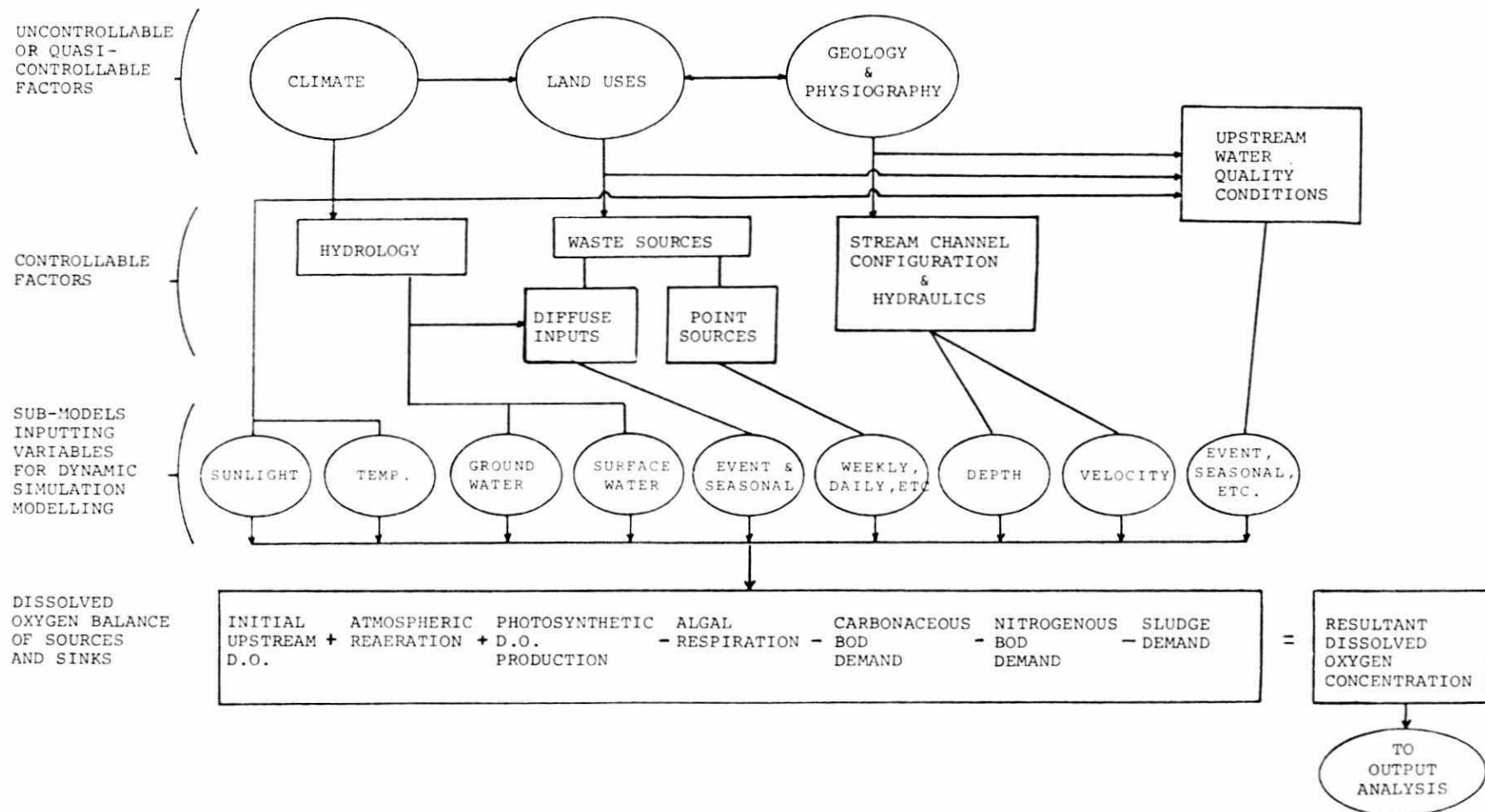
long-term water quality and quantity data and data collected during field investigations.

There are several ways in which the simulation can out put information about water quality at any specified location in the basin. To identify an "event" (i.e. dissolved oxygen level depression) a reference point or standard must be defined. For example, the Ontario Ministry of the Environment's 5 mg/l minimum dissolved oxygen criterion for protection of warm water biota may be selected as the standard. The model can then be used to predict the probability of violating that standard. The "event" can be described in several dimensions:

  a. the probability of violations occurring in a specified period;

  b. the magnitude of violations; and/or,

  c. the duration of violations.

In employing the dynamic simulation for developing water resource management plans, a number of options for many of the inputs to the model can be incorporated. For example, the level or type of wastewater can be varied (including zero pollutant). Improved upstream water quality conditions and streamflow augmentation can be included. Wastewater outfall relocation, channel improvements and other physical alterations can be considered.

The foregoing is a general overview of dynamic simulation modelling, illustrating the flexibilities and capabilities of this valuable new water resource assessment planning tool. To restate the basic advantage of this modelling technique over others employed: simulation permits the incorporation of variability of modelling components and with this capability can predict daily, weekly or seasonal water resource problems that might otherwise be overlooked.

RECENT IMPROVEMENTS IN ASSESSMENT PROCEDURES

Waste load variability, photosynthesis and respiration, and nitrification have been incorporated into water resource assessment techniques over the past few years. In many areas of the province, they have pronounced impact on water quality and use.

1. <u>Waste Load Variability</u>

A detailed evaluation of the variability of BOD_5 and other components in a wastewater discharge is absolutely essential as the water quality of streams is closely related to the quality and quantity of waste discharge.

Despite being designed to achieve a specified level of treatment, it is generally recognized that treatment plants do not discharge a constant quality and quantity of effluent day-by-day or even hour-by-hour. At most plants there is considerable variation in effluent quality.

As an example, effluent samples from a secondary sewage treatment plant generally typical of most Ontario plants varied in quality from 4 to 55 mg/l of BOD_5, or in terms of daily loading incorporating plant flow rates, ranged from 380 to 7450 pounds of BOD (Table 1).

To illustrate potential errors in modelling, it would be assumed that the effluent BOD concentration from an activated sludge plant is 15 mg/l and from this figure it would be determined that the average daily BOD load is about 2100 pounds. In fact, the actual daily loading exceeds this average value about one quarter of the time.

In a river system where waste loading guidelines are designed to meet minimum designed dissolved oxygen levels based on average or design effluent quality, waste loading fluctuations above the mean could have devastating effects on downstream quality and use. In the past, such occurrences were hopefully avoided by applying a safety factor ranging from 20 to 40 percent; that is, only 60 to 80 percent of the stream's waste loading capacity was allocated to waste sources. The remaining capacity was held in reserve in part, to protect the watercourse in case of plant upsets.

At this time, it is not common practise to develop a non-steady state simulation model of sewage treatment plant operation. Therefore, plant discharge sub-models presently accepted by the simulations are based on statistical analyses of effluent data. A good historical record of plant effluent quality and quantity data is therefore desirable. In those cases where an effluent quality to flow relationship cannot be established, a probablistic distribution can be defined based on the cumulative distribution functions of the data. These statistical relation-

TABLE 1: SEWAGE FLOW AND BOD₅ VARIATION AT A SOUTHERN ONTARIO

SECONDARY SEWAGE TREATMENT PLANT

FLOW (MGD)	BOD₅ MG/L	POUNDS/DAY	FLOW (MGD)	MG/L	POUNDS/DAY
11.8	6	710	9.1	21	1900
11.8	7	770	10.1	20	2020
12.3	8	900	10.1	14	1410
12.7	9	1140	10.2	8	780
11.4	55	6270	15.0	8	1170
12.1	40	4840	10.6	10	1060
9.9	4	400	10.1	10	1010
11.6	24	2780	11.0	33	3630
11.5	12	1380	8.7	17	1470
11.6	12	1390	11.1	11	1220
10.8	20	2160	11.3	15	1700
10.0	13	1300	9.8	11	1070
10.7	7	750	9.9	20	1980
11.9	10	1130	9.9	20	1970
10.1	50	5050	8.9	20	1780
10.3	26	2680	11.5	10	1150
9.8	19	1860	8.0	7	590
11.0	21	2310	9.9	9	870
9.4	47	4420	11.1	12	1330
9.5	5	430	8.9	28	2480
11.6	6	700	10.1	9	890
12.1	24	2910	10.1	5	530
10.5	32	3360	13.0	12	1560
13.3	14	1860	11.8	10	1180
9.7	4	390	9.7	23	2230
8.8	21	1850	9.9	21	2080
9.9	17	1680	8.1	23	1860
10.0	9	880	10.1	32	3220
10.8	12	1300	14.1	53	7450
11.7	24	2810	9.1	22	2010
9.1	6	580	9.0	11	990
12.3	10	1180	8.4	9	740
11.5	15	1720	8.3	15	1250
12.4	11	1360	9.9	13	1290
10.2	8	860	10.7	20	2150
9.6	4	380	12.5	52	6480

ships appear to work quite well if the regression analysis is not used for extrapolation beyond the range of the data used in its development.

2. Photosynthesis and Respiration

In both the steady-state and dynamic simulation models, photosynthesis and respiration factors are included. These processes can have a significant influence on the dissolved oxygen levels in streams, ponds and lakes and are directly related to aquatic plant and algae growth. Photosynthesis refers to oxygen produced during the photosynthetic process of aquatic plants during hours of daylight. "Respiration" is the term applied to the daily consumption of dissolved oxygen by plants.

The daily production-consumption cycle of oxygen at any location in a watercourse is a function of the aquatic floral biomass present and sunlight intensity. Weed and algae growth are in turn controlled by physical factors such as substrate, river velocity, water temperature, turbidity and the availability of macro and micro nutrients.

Photosynthesis and respiration can markedly affect the dissolved oxygen regime of a watercourse and severely restrict use. It is not uncommon in a productive shallow stream to measure day-time and night-time fluctuations of dissolved oxygen from 200 percent saturation to 20 percent saturation. Occasionally, complete oxygen depletion occurs during the early hours of the morning. Figure 2 shows diurnal fluctuations of dissolved oxygen in a very productive stream in southwestern Ontario. This is not atypical of many other algae-choked streams in southern Ontario.

The effects of photosynthesis and respiration can readily be incorporated into either the steady-state models or simulations and appropriate rates for a given level of biomass and sunlight intensity can be measured in the field using continuous recording dissolved oxygen meters.

Biomass production is a function of physical conditions and the availability of key elements such as the nutrient phosphorous. Difficulties arise when photosynthesis and respiration terms must be predicted because biomass growth levels must also be predicted. In these cases, when using

Figure 2

OBSERVED AND PREDICTED DIURNAL DISSOLVED OXYGEN PROFILE

Middle Maitland River near Listowel - August 30,1972

⊙ - Observed
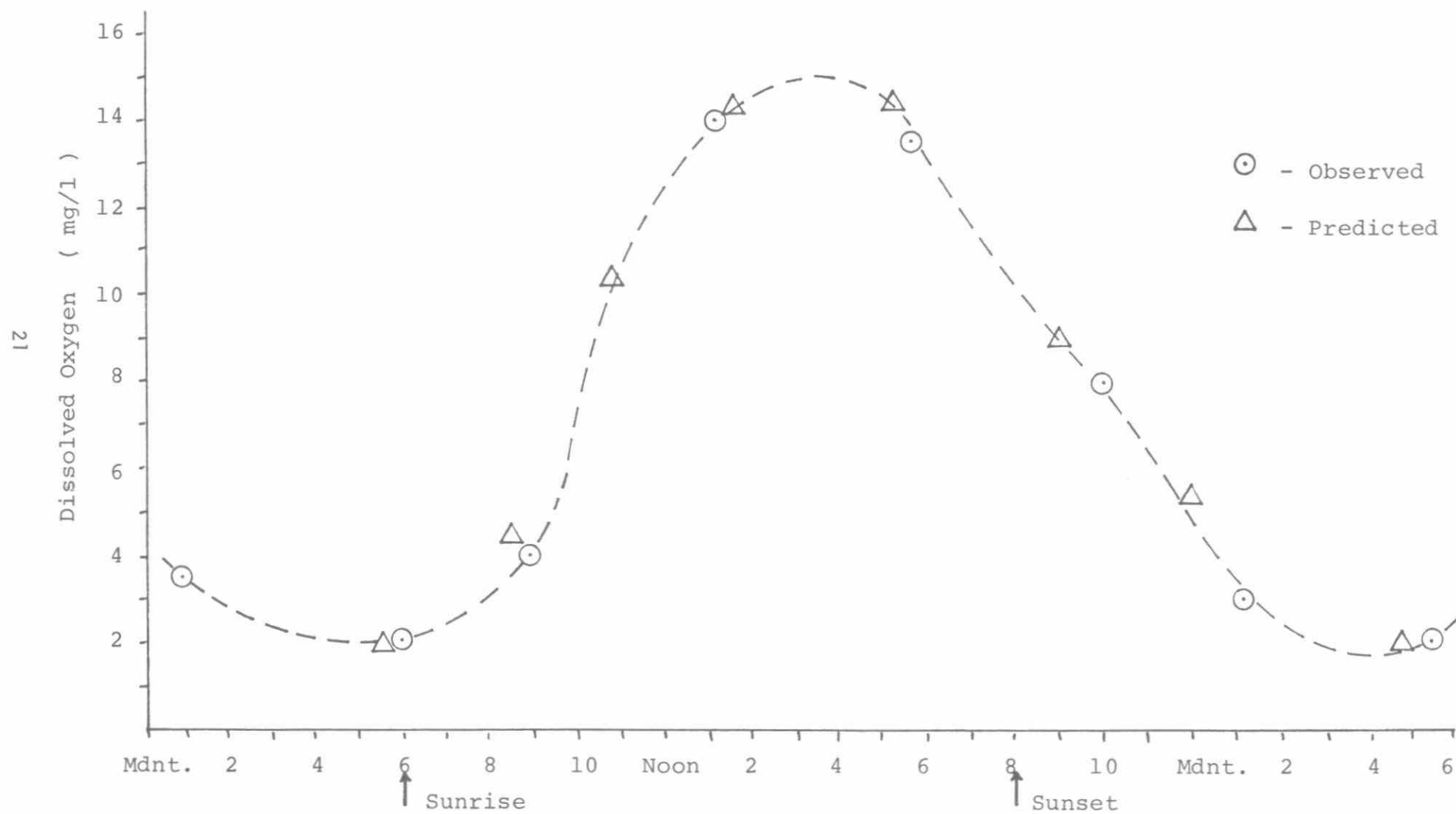
△ - Predicted

steady-state models, the highest level of biomass production in critical sunlight intensity conditions for the appropriate period of the year are employed. Predictive capabilities are improved in the simulation model because this system builds in the flexibility of altering the daily sunlight intensity conditions by selecting this term at random from a given range. A biomass level, however, must still be arbitrarily selected.

Studies are currently underway by Ministry of the Environment staff to develop relationships between nutrient (i.e. phosphorus) levels in the watercourse and biomass production. When this relationship is established, models will be available to predict biomass levels (and resulting DO fluctuations) for any given nutrient loading condition, or more importantly this relationship will permit the calculation of allowable phosphorus loadings from various waste sources that will result in a pre-determined level of biomass production. The level of production itself would be based on maximum allowable diurnal dissolved oxygen fluctuations or aesthetics.

3. Nitrification

The discharge of nitrogenous compounds into watercourses produces a variety of changes in water quality. Changes not only occur in the various forms of nitrogen but also in those substances with which nitrogen reacts. The utilization of dissolved oxygen for the conversion of organic nitrogen (ammonia) to the stable nitrate form, the utilization of nitrogen in its various forms as a nutrient by aquatic plants and algae, and ammonia interference in wastewater chlorination are significant examples.

Organic and inorganic nitrogenous compounds are found in municipal wastewaters, some industrial effluents, urban storm drainage and agricultural runoff. In some areas the natural input of nitrogen compounds from decaying vegetation etc., is significant.

In water resource modelling, one of the most significant impacts of nitrogen compounds is the demand on oxygen resources. Nitrogen introduced to a watercourse as ammonia is oxidized under aerobic conditions by bacterial action to nitrite. This action requires 3.43 grams of oxygen for each gram of ammonia nitrogen converted. Nitrite in turn is oxidized to the stable inorganic nitrate nitrogen. This process requires 1.14

grams of oxygen for each gram of nitrite converted. The total oxygen utilization in the conversion of 1 gram of ammonia to nitrate is therefore 4.57 grams.

Until recently, the nitrogen component of the ultimate bio-chemical oxygen demand has not been incorporated into water quality assessments. The nitrogenous portion of this total demand from a secondary effluent is seldom reflected in the standard 5-day BOD test.

Reasons for this include the fact that the carbonaceous component of the oxidizable materials are satisfied first, usually within the first four to seven days. Secondly, nitrifying bacteria required for the conversion of ammonia to nitrate are either not present in sufficient numbers or are suppressed in treated effluent. The nitrogen component, however, does appear to dominate about the sixth or seventh day and is the largest component of the total oxygen demand from that point until oxygen demand is totally satisfied - usually about 20 to 25 days.

A laboratory BOD satisfaction curve showing the carbonaceous and nitrogenous components is shown in Figure 3.

In natural watercourses, the nitrogenous oxygen demand usually becomes a factor much more quickly because the nitrifying bacteria, inhab-itants of the soil, are usually available in high numbers. It is quite obvious therefore that the potential oxygen demand of nitrogen compounds in waste discharges should be determined so that correlation can be established between the potential BOD measured in a waste discharge and the actual dissolved oxygen utlization as measured directly in the receiving watercourse.

Nitrogenous oxygen demand can be measured directly by a 20-day BOD test or a nitrogenous BOD test. It can also be approximated by evaluating the Kjeldahl, ammonia and nitrate concentrations in a wastewater discharge.

In many southern Ontario River basins currently approaching, or in a few cases, exceeding their oxygen consuming waste receiving capacity, nitrification (i.e. conversion of organic nitrogen to the nitrate form) can be adequately accomplished by extending retention time in the aeration tanks and is being specified for many new plants or plant expansions.

Figure **3**

CARBONACEOUS AND NITROGENOUS COMPONENTS

OF THE ULTIMATE BIOCHEMICAL OXYGEN DEMAND
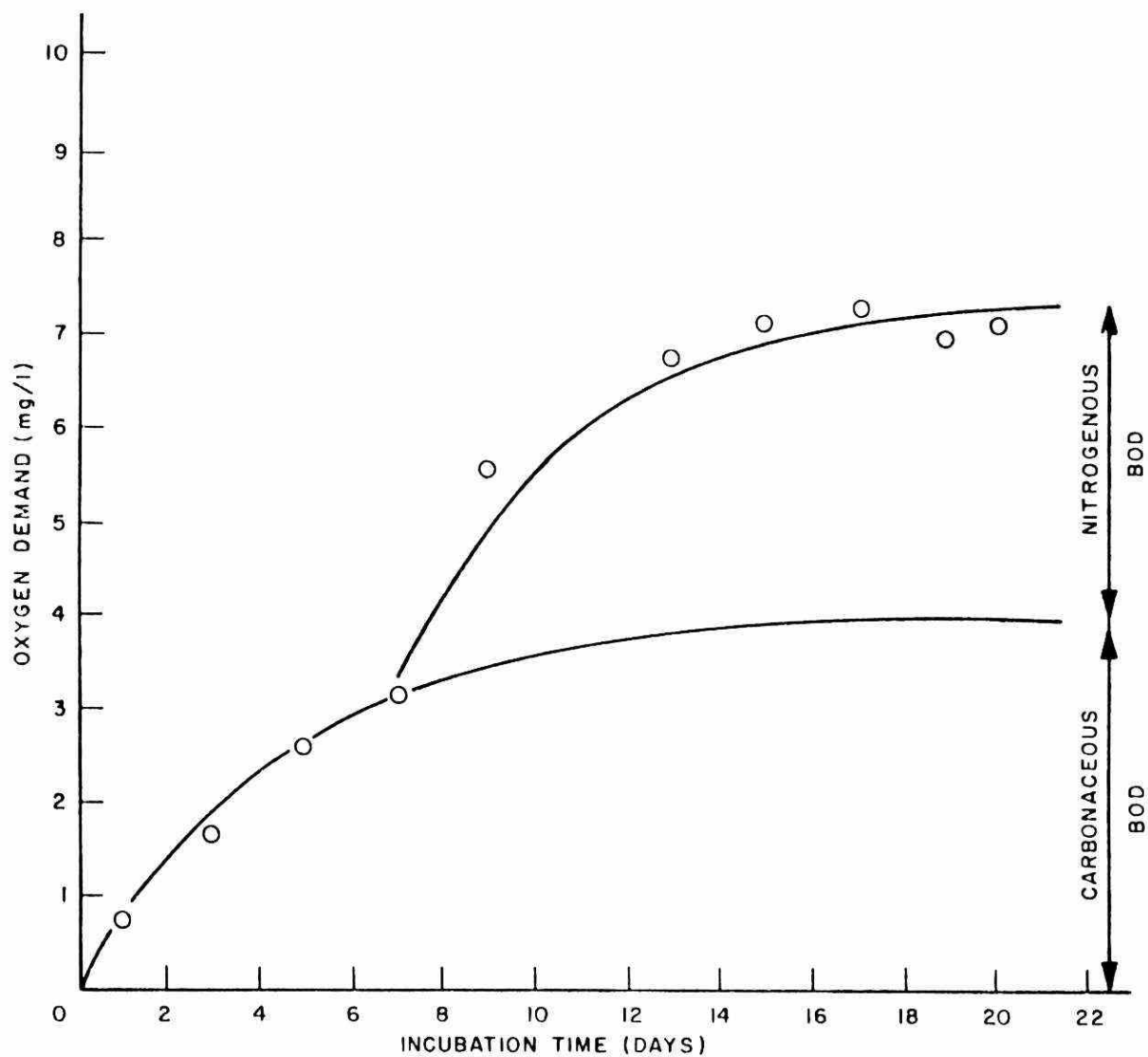
Nitrogen control and sewage treatment works is being discussed in detail in later presentations.

4. Other Recent Advances In Water Resource Assessments

Simulation modelling, diurnal dissolved oxygen fluctuations caused by photosynthesis and respiration, and nitrification are three areas of water resources assessment technique development which have been incorporated into practicable application by the Ministry of the Environment over the past few years. A number of other areas are currently under investigation for future application. Very briefly these subjects include,

a. Environmental Impact of Sewage Disinfection.

The use of chlorine to disinfect wastewater and protect the receiving waters and downstream users from bacterial contamination is common practice in Ontario. Chlorine is a very toxic substance and chlorine residuals in watercourses downstream from sewage treatment plants can damage or eradicate fish and other aquatic life in their stream or lake.

While public health protection from bacterial contamination remains a prime concern, chlorine residual levels that will not impair aquatic life are being studied. The impacts of alternatives to chlorination such as ozonation are also being investigated. Ozonation offers a side benefit of adding dissolved oxygen to the wastewater discharge and all sources of oxygen in the polluted stream are beneficial; however, other undesirable factors both in the plant and receiving waters may exist and render ozonation unacceptable.

b. Urban Land Runoff

Recent studies undertaken in Ontario and elsewhere have shown that the quality of urban storm water runoff can at times be poorer than raw municipal sewage. As well as containing high levels of BOD and suspended solids, storm drainage contributes nutrients and heavy metals to the receiving watercourse.

The aquatic environmental impact of urban storm drainage is difficult to assess because of the relatively short duration of the event. Certainly the materials introduced during a storm contribute to the long-term build-up of pollutants in a river or lake and recent studies have

indicated that storm drainage is resulting in measurable degradation down-
stream, during and shortly after the event.

Further studies of the characteristics of urban storm drainage,
including bypassed municipal wastes and their impact on water resources,
may lead ultimately to storm waste collection, storage and treatment.

THAMES RIVER BASIN WATER MANAGEMENT STUDY

by

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INTRODUCTION

This paper presents some of the findings of the Thames River
basin water management study, undertaken jointly by the Ontario Ministries
of the Environment and Natural Resources. The study was initiated in 1972
in response to growing concern over existing problems relating to water
quality, flooding and erosion in the watershed, and over potential pro-
blems anticipated as a result of future population growth and economic
development. The complexities and interrelated nature of the problems
emphasized the need for a water management plan encompassing the entire
drainage basin.

The Thames River basin is the second largest in Southwestern
Ontario, and drains an area of approximately 2,250 square miles. Its total
length from the source of the North Thames River to Lake St. Clair is
approximately 125 miles. Major water uses in the basin include water supply
for agricultural, domestic, municipal and industrial purposes; waste
disposal and assimilation; recreation, and fish and wildlife habitat.
Inherent conflicts among these uses are prevalent in the basin. Moreover,
proposed solutions to the problems may themselves create additional con-
flicts.

WATER RESOURCES PROBLEMS

The two main water management problems in the Thames River
basin are water quality impairment and flooding. Impairment of surface
water quality is primarily caused by excessive inputs of nutrients, oxygen
consuming materials, bacteria and suspended solids. Major urban sources of
these contaminants include sewage treatment plant effluents, and storm and
combined sewer discharges. Municipal drains, field tile systems, surface
runoff from fertilized fields, drainage from intensive feedlots, treated
effluent from rural industries, and the free access of cattle to streams

are major rural sources of water quality impairment. Excessive aquatic plant growth and unpleasant aesthetic conditions are the most visible signs of water quality impairment; however, the less visible problems of low dissolved oxygen levels and high bacteria levels are also significant. This impairment has led to the curtailment or restriction of legitimate water uses in the watershed. Most severely affected by this impairment are fish and aquatic life and recreational water uses.

Recurrent flooding is the other most significant problem in the watershed, particularly in St. Marys, Woodstock, London, and the area from Thamesville through Chatham to Lake St. Clair. Average annual flood damages in the watershed were calculated to be over 1.5 million 1975 dollars, of which 57 percent is in Chatham and 20 percent in the vicinity of London. Related in part to flooding is erosion of streambanks and dikes, primarily in the lower watershed. Erosion of topsoil is also a significant problem.

The inadequacy of water-based recreational facilities to meet demands and the potential loss of prime agricultural land were also identified as problems common to the watershed. Other management problems of local importance include negative effects of artificial land drainage, water supply interference and ground water quality impairment. Communication and co-ordination problems were also noted. For a more complete discussion of these findings, reference should be made to the Thames River Basin report.

WATER MANAGEMENT OBJECTIVES

Since environmental and study objectives are fundamental to the selection of effective water management proposals, an outline of these objectives is warranted. The overall objective of the study is:

> "To develop guidelines for water management planning in the Thames River basin which would ensure that an adequate quantity of water at a satisfactory quality is provided for the recognized water uses in the river basin at the lowest cost and that flood and erosion protection is provided consistent with appropriate benefit-cost criteria."

As indicated earlier, stream water quality is the major water resource problem in the basin. Water quality improvement depends upon the identification of appropriate water quality objectives and the implementation of courses of action selected to achieve these objectives.

The basic philosophy of the Ministry of the Environment is that there should be constant effort to improve water quality, recognizing that improving the quality of water makes it available for more uses. However, there are certain minimum levels of water quality generally acceptable to the province which must be met. In this regard, water quality of requirements established for the protection of aquatic life are normally selected. Higher levels of quality are required in areas where more demanding uses such as swimming and bathing occur.

Maintaining areas of presently high quality waters and upgrading quality elsewhere to a level which would protect fish and aquatic life was taken as a realistic and obtainable water quality objective for purposes of this study. However, it must be stressed that the long term objective of the Ministry of the Environment is to upgrade water quality in the basin as much as possible.

Having established the objective of maintaining water quality necessary for the protection of fish and aquatic life, appropriate criteria to achieve this objective and to protect other uses, such as recreation and aesthetics, were also identified. A basin-wide study was carried out into fish distribution and corresponding dissolved oxygen requirements. It was estimated, in probabilistic terms, what the minimum dissolved oxygen concentrations should be on a daily, monthly and seasonal basis to maintain a viable fishery. Specific dissolved oxygen objectives were defined by application of criteria to specific stream reaches. Other water quality criteria applicable to fish and aquatic life, recreation and other water uses are specified in the publication "Guidelines and Criteria for Water Quality Management in Ontario" (Ministry of the Environment, July, 1974).

The objective of flood control activity is to minimize the average annual flood damage with the least cost. The primary constraint on this evaluation is that the average annual flood damage reduction must exceed the average annual cost of achieving that reduction. The cost in this respect must reflect the actual costs of constructing a flood control facility, and the conflict costs which can be allocated to any interference that the flood control facility may have with other uses or any other part of the system. This cost figure is reduced appropriately by other benefits which accrue, such as flow augmentation and recreation.

The main emphasis in the planning of water control facilities in the Thames River Basin to date has been on flood protection. However, the final design and operations have usually incorporated and developed the multiple purposes of flood control, water supply, pollution abatement and recreation. The emphasis in this study provided for the integration of all water management activities within the watershed to reflect the optimum system and operating policy.

MAJOR WATER MANAGEMENT OPTIONS

Methodology

In order to develop effective and realistic water management guidelines, the study involved an assessment of the availability and quality of both surface and ground water, an inventory of water uses and related land uses, and an evaluation of existing and potential water resource problems in the basin.

One feature of the study was the intensive use of mathematical modelling in the evaluation of waste treatment and reservoir alternatives with regard to the effect on water quality parameters and flood control benefits.

Another feature of the study was the Public Consultation Program (PCP), designed to provide municipal officials and the residents of the basin with an opportunity to express their views concerning the management of their water resources.

Major options which have implications for the greater portion of the basin in terms of water quality improvement and flood control, the two primary objectives of the study, are the construction and operation of dams for flow augmentation, flood control and recreational use and the sewage disposal options for the City of London. The primary evaluation criteria used in evaluating these options was the flood control benefit-cost ratio and total system net cost in present value terms, which took into account as much as possible this interrelationship of flood control and water quality benefits. For example, construction of a dam provides economic benefits in terms of both flood control and water quality, the latter accruing from deferral of capital expenditures for sewage treatment facilities downstream from the dam due to flow augmentation provided from the reservoir.

Some of the major options evaluated could be immediately dis-
missed for one of two reasons:

1. They totally failed to meet one or both of the required
   primary objectives of improving water quality and increas-
   ing flood control; or,

2. They placed severe growth restrictions upon the City of
   London.

The reason for discarding options falling into category 2
above requires some explanation. The approach taken toward the overall
water resource management of the basin was to optimize water resource use.
Accordingly, in areas where the limit to the capacity of resources to
sustain growth is reached within the planning horizon, a policy of growth
limitations may be appropriate with excessive growth redirected to desig-
nated centres, primarily the City of London. Provincial planning studies
have recognized London to be the major growth centre of this region and
have recommended that it continue in that role. Thus, only those options
which would allow London to expand to its 2001 project population of
500,000 were considered. However, it is important to note that a signifi-
cantly lower population growth rate would fundamentally affect the
evaluation of waste treatment options at London.

Once the major option configurations had been determined and
evaluated on the total system cost basis, secondary evaluation of major
options began. At this stage, as yet unquantified parameters, as well as
those factors which cannot be quantified were considered. These include
the objective of minimizing the loss of prime agricultural land and
environmental disturbances due to capital construction projects, especially
dams, and of increasing water based recreational facilities. The results
of the Public Consultation Program provided considerable support in this
procedure by removing some of the subjectivity from the evaluation process.
An imputed value for non-quantifiable factors can be derived, however.
If the decision maker chooses an option which is not least-cost on the
basis of the total system cost analysis, then the added cost of the option
chosen should be equal to or less than the net unquantified benefits
which the chosen option offers over the least-cost option.

At the primary evaluation stage, capital construction costs
and flood control benefits have been used in deriving total net costs

for each system option. In simplified terms this consists of estimating
the capital costs for construction of engineering works (dams, pipelines,
treatment plants) and the year in which they are to be constructed. The
total cost in present value terms is then calculated using various dis-
count rates - 2, 4, and 7 percent. Flood control benefits in similar pre-
sent value terms are then subtracted as they represent negative costs, to
produce the total net cost for each system option. Costs are presented
in present value terms so that they can be compared at a single point in
time. The present value takes account of the time when costs and benefits
occur by weighting near-term dollars more heavily than those far off in
the future.

The five dams which have been proposed in the past and the sewage
treatment options for London are discussed below both individually and in
system configurations such that all associated benefits and costs, economic,
environmental and social, are brought to light. Preliminary screening of
several hundred theoretical combinations reduced the number of options
to be evaluated in detail to 22. For brevity of presentation, all planning
options in this section are presented in compact notation form as follows:

Gg: Construct Glengowan dam and reservoir
       primarily for flood control;

Gg*:   Construct Glengowan dam and reservoir
       primarily for flow augmentation;

Th: Construct Thamesford dam and reservoir;

W: Construct Wardsville retarding structure;

CC: Construct Cedar Creek dam and reservoir;

ZS: Construct Zorra Swamp dam and reservoir;

P: Construct a sewage trunk pipeline from
       London to Lake Erie;

T: Provide tertiary sewage treatment for the
       City of London in order to meet effluent
       requirements.

In all cases, dam construction is assumed to be completed in
1981. For "T" and "P", the number following the notation refers to
the year when that option would be operational (Table 2). A system
option is designated by a combination of two or more single options.

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Thus, for example, option Gg + Th + P:94 indicates that the Glengowan and Thamesford dams would be constructed to be operational in 1981 and the London-Lake Erie pipeline and accompanying treatment plant would be constructed to be operational in 1994.

The waste management alternatives available to the City of London in order to meet and maintain water quality objectives in future, can be reduced to the following options:

1. Implement tertiary treatment (discharging an effluent of approximately stream quality) and continue discharging to the Thames River.

2. Export sewage via pipeline to Lake Erie for secondary treatment and phosphorus removal prior to discharge.

3. Build and operate Glengowan reservoir primarily for flow augmentation, and continue discharging to the Thames River using conventional treatment providing an effluent equivalent in quality to the Greenway plant.

An economic analysis of these options similar to that for flood control options is not possible. The benefit of water quality improvement or maintenance cannot be readily quantified, unlike flood control benefits, and hence, benefit-cost analysis is not possible. Minimum water quality criteria must be met, however. Given this objective, economic analysis is applied to determine the least-cost method of achieving it. As previously explained, the costs of sewage treatment alternatives for London are related to upstream reservoir construction options and must therefore be evaluated in a total system context.

Tertiary treatment as discussed here is taken to include traditional secondary sewage treatment plus the following processes:

- phosphorus removal;
- carbon adsorption;
- filtration; and,
- ammonia stripping.

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Option Analysis

From results of the water quality computer simulation model developed for this study, waste loading guidelines for the City of London were generated (Table 1). The total allowable load (total oxygen demand) varies with the amount of river flow. Hence, the allowable loadings were increased for those options which include flow augmentation from Glengowan, Thamesford, or both.

In addition to providing total load guidelines, specific limitations have also been placed on discharges to the North Thames and the south branch of the Thames within the city. Furthermore, loading figures apply only to sewage treatment plant effluents and are based upon the assumption that effects of urban runoff do not increase with time. This can only be achieved by control of discharges from combined sewer overflows and, as population increases, by storm water treatment.

Table 1 also shows dilution ratios (streamflow/sewage flow) which can be expected to occur under low flow conditions. This factor is important since the effects of pollutants which have not been modelled, such as heavy metals and other toxicants, are not precisely known and must also be controlled. The current dilution ratio in London is approximately 1.5:1. With increased levels in treatment, this ratio may be allowed to decrease. However, the ratio should not be allowed to decrease below 1:1, at least until the effects of this flow ratio upon water quality and fishlife have been determined. For that reason, implementation of additional remedial action is required by the dates at which this dilution ratio is reached.

If the population figure of 500,000 for London is accepted as the maximum population for the city as has been suggested, then it can be seen from Table 1 that, with an allowable dilution ratio of 1:1 and construction of one additional dam upstream, tertiary treatment is a viable option. An advantage of this option is that it completely avoids any environmental effects that export of sewage might have upon Lake Erie and any pipeline right-of-way.

Table 2 presents the net cost and least cost order of the various systems options evaluated in this manner.

TABLE 1:   LONDON WASTE LOADING GUIDELINES

Option	Total Allowable Load Oxygen Demand lb/day			Year Load Limitation Reached[1]	Year Dilution Ratio Reached	
	Total	N. Thames	S. Branch		1.5:1	1:1
Present Conditions (no additional flow augmentation)	8,000	1,000	2,500	1993	1971	1984
Gg	11,000	2,000	2,500	2001+	1983	1997
Gg*	17,000	4,000	2,500	2001+	1999	2001+
Th	11,500	1,000	3,500	2001+	1986	2001
Gg + Th	14,500	2,000	3,500	2001+	1994	2001+
Gg* + Th	21,000	4,000	3,500	2001+	2001+	2001+

[1] At treatment to stream quality.

Gg* - Glengowan operated primarily for flow augmentation.

TABLE 2: TOTAL SYSTEM NET COSTS

Option	Net Cost Present Value ($M)			Least Cost Order		
	2%	4%	7%	2%	4%	7%
1. W+T:81	95.4	89.4	77.2	19	19	22
2. W+Gg+T:83	101.1	91.5	75.8	21	22	20
3. W+Gg+P:83	82.1	71.2	62.6	16	14	16
4. W+P:81	75.2	71.5	61.8	8	15	15
5. W+Th+T:86	83.3	78.7	66.6	17	18	18
6. W+Th+P:86	67.7	65.0	55.2	5	11	12
7. Th+T:86	80.7	74.2	61.4	14	16	14
8. Th+P:86	65.2	60.5	49.9	4	7	10
9. Th+Gg+T:94	76.6	63.7	47.4	11	9	8
10. Th+Gg+P:94	71.1	58.8	43.5	7	5	6
11. Gg+T:83	105.3	91.1	72.7	22	21	20
12. Gg+P:83	86.3	74.7	59.5	18	17	13
13. W+Gg+Th+T:94	81.2	69.3	53.5	15	13	11
14. W+Gg+Th+P:94	75.3	64.4	49.3	9	10	9
15. W+Gg*+P:99	77.3	60.2	41.3	12	6	4
16. W+Gg*+Conventional Treatment**	23.1	27.0	26.2	3	3	2
17. Th+Gg*+Conventional Treatment	16.6	22.0	22.5	1	1	1
18. W+CC+ZS+T:82	98.9	90.7	76.2	20	20	21
19. W+CC+ZS+P:82	80.0	67.5	63.2	13	12	17
20. Th+CC+ZS+T:93	75.8	63.1	46.4	10	8	7
21. Th+CC+ZS+P:93	70.3	58.2	42.5	6	4	5
22. Th+CC+ZS+ Conventional Treatment	20.1	26.1	26.8	2	2	3

* Involves operation of Glengowan dam primarily for flow augmentation.
** Option 16 involves growth limitation of 480,000 at London.

Note: W, Gg and Th - assume construction in 1981.

As well as taking into account flood control benefits, the figures in Table 2 include benefits (negative costs) to London attributable to the flow augmentation, which defers capital construction of a pipeline "P" or tertiary sewage treatment plants "T". For example, with no upstream dam construction as in option 1, "T" is required in 1981; with one upstream dam as in option 2, "T" is required in 1983; and with two upstream dams as in option 13, "T" is not required until 1994. The net costs of options 18 to 22 exclude both the economic benefits of flow augmentation to Woodstock in terms of deferred treatment expenditures and the increased water supply costs resulting from flooding of the Woodstock well field.

It should be noted that the economic analysis undertaken to arrive at the total system least-cost ordering is extremely sensitive to population projections and associated sewage flows and hence, to deferral times of capital expenditures. For example, it can be seen from Table 2 for options 3 and 4 that the added expenditure of "Gg" construction in option 3 is totally balanced out by a two year deferral of "P" construction. Hence, system options 3 and 4 emerge as having an almost identical "net cost". In view of this, the phase-timing of each option, and especially of "P" and "T" should be carefully noted as any change in this timing will significantly affect the net cost of the option concerned.

Given the primary objectives of achieving good water quality throughout the watershed at the lowest cost and providing flood control consistent with benefit-cost criteria, the three unquantifiable objectives mentioned earlier were taken into consideration in this evaluation of the main least-cost options.

Conclusion

When all these factors had been considered, it was concluded that the preferred option is to construct the Thamesford dam primarily for flood control, the Glengowan dam primarily for flow augmentation, and to utilize conventional treatment at London. However, if it is decided that development of a limestone deposit precludes construction of the Thamesford dam, the preferred option is to construct the Wardsville dam for flood control, the Glengowan dam primarily for flow augmentation, and to utilize conventional treatment at London. If the growth limitation of

TABLE 3.   SUMMARY OF UNQUANTIFIED COSTS AND BENEFITS OF MAJOR OPTIONS

Option	Unquantified Costs	Unquantified Benefits
Glengowan Dam	1. Agricultural land permanently inundated. 2. Water quality deterioration in and immediately downstream from the reservoir. 3. Increased water temperatures. 4. Disruption and destruction of some fish and wildlife habitats.	1. Improved recreational opportunities, either directly through provision of facilities at Glengowan reservoir or indirectly through improved water quality in the Fanshawe reservoir. 2. Improved water quality through flow augmentation in downstream areas. Flow augmentation benefits to London have been included in the economic analysis. 3. Flood control benefits, not included in the economic analysis, if Glengowan is used primarily for flow augmentation.
Thamesford Dam	1. Agricultural land permanently inundated. 2. In-reservoir water quality deterioration. 3. Water temperature increased. 4. Sports fishery disruption. 5. Some disruption and destruction of fish and wildlife habitats. 6. Public opposition voiced. 7. Foregone opportunity to extract limestone deposits.	1. Possible recreation benefits if facilities provided. 2. Water quality improvement through flow augmentation to downstream areas. This benefit to London was included in the economic analysis.
Wardsville Retarding Dam	1. Disruption of road links between Elgin and Middlesex countries during high flows, possibly requiring new bridge construction. 2. Strong public opposition expressed. 3. Occasional inundation of Indian Reserve lands.	

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TABLE 3.   (CONT'D)

Option	Unquantified Costs	Unquantified Benefits
Cedar Creek Dam	1. Inundation of the Woodstock well field. 2. Disruption of a stocked and natural coldwater fishery. 3. Destruction of a major deer-yarding area. 4. Agricultural land permanently inundated. 5. In-reservoir and downstream water quality impairment. 6. No recreational benefits.	1. Additional dilution flow for Woodstock and downstream areas through low flow augmentation.
Pipeline	1. Environmental effect upon Lake Erie and pipeline corridor, and pressure for urban development along the corridor. 2. Reduction of flow in the Thames River.	1. Possible advantages to muncipalities along the pipeline corridor outside the watershed should they tie in to the system. 2. Removal of all London sewage effluent from the river giving improved water quality downstream. 3. Possible benefit of direction of controlled growth away from more sensitive areas.
Tertiary treatment	1. Eventual growth restraints on London City due to dilution ratio limitations.	1. Allows for any advantages future sewage treatment technology improvements may offer.
Zorra Swamp	1. In-reservoir and downstream water quality impairment. 2. No recreational benefits. 3. Major changes in the natural ecology of the swamp.	1. Additional dilution flow for downstream areas through low flow augmentation.

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480,000 for London associated with this option is decided to be unacceptable, then other options, such as provision of tertiary treatment or construction of sewage pipeline to Lake Erie can be considered.

As the Glengowan dam is common to each of the preferred options, construction of the Glengowan dam first would offer maximum flexiblity in choosing other capital projects. Decisions as to whether to construct the Wardsville dam or the Thamesford dam could then be made. The decision as to whether to utilize conventional treatment or eventually a sewage pipeline from London to Lake Erie could be deferred to the early 1990's.

LOCAL WATER MANAGEMENT OPTIONS

Introduction

Although the major options have great significance to basin wide water management, they by no means deal with all the basin's water resource problems. Local water management problems can have a cumulative effect, so that a localized type of problem, recurring at several different locations, can have basin wide implications. A wide range of management options to deal with urban, rural, reservoir related and flooding problems has been considered and applied on a stream reach and municipality basis.

Urban oriented options include varying levels of treatment of sewage and industrial wastes, and growth restrictions. In areas where the remaining waste assimilative capacity of streams is limited, municipalities proposing additional growth can consider the installation of advanced tertiary waste treatment plants as defined earlier, producing a highly polished effluent equivalent to stream water quality, or waste storage for summer spray irrigation or discharge during periods of adequate flow. However, for smaller municipalities, the costs of the required tertiary treatment may be prohibitive. Moreover, the costs of property acquisition for waste storage can make this uneconomical and this approach often involves the use of prime agricultural land. The alternative to the above treatment options is growth restrictions.

Receiving streams at other municipalities in the basin have varying capacities to assimilate additional waste loadings. The additional assimilative capacity at some municipalities is limited and long term growth would be inadvisable from a water quality viewpoint. At others, the

additional waste assimilative capacity is not as limited.

Control of urban runoff is an important consideration in the basin. Although the significance of pollution loads from this source at each municipality was not documented during this study, urban runoff is recognized as a source of stream impairment. Rural oriented management practices for water quality improvement include limiting fertilizer application rates, channel protection programs, restricting free access of cattle to streams, control of farm waste discharges, particularly from intensive feedlot operations, and control of illegal septic tank connections to drains. Surface runoff to streams from fertilized land is a significant diffuse source of nutrients which contributes to excessive aquatic weed growth. Although accurate statistical information is not available, fertilization of cropland beyond recommended rates was found to be a general practice in the basin. There are other localized problems such as channel and soil erosion or flood plain development controls, which while important, are somewhat outside the scope of this Seminar. Anyone interested in more details is referred again to the Main Thames River report.

The following case studies taken from the Thames River basin report illustrate the use of these control options in solving individual municipal pollution problems.

Stratford

Population projections for the City of Stratford indicate that a population of 47,000 could be expected by the year 2001. Water quality studies have shown that even at the present population and under a streamflow regime 40 times higher than the estimated low flow, minimum dissolved oxygen levels of 2 mg/l severely violate criteria for the entire length of the stream. Although municipal wastes from Stratford receive advanced treatment to reduce organic loadings, consisting of secondary treatment with effluent filtration, modelling studies have shown that oxygen guidelines in the entire Avon would not be achieved even with the total elimination of oxygen demanding materials from Stratford. This situation is the result of obnoxious aquatic plant growth caused by a continuous nutrient availability, coupled with extremely low natural streamflows during summer months. As previously stated, it is the long term objective of the Ministry of the

Environment to upgrade water quality to make the resource available for broader use. Management approaches directed towards the achievement of dissolved oxygen objectives set for the immediate objective of sustaining fishlife in the Avon River are: harvesting of aquatic vegetation from the receiver; advanced phosphorus removal during aquatic vegetation growth periods; further treatment to remove nitrogenous organics; storm water treatment; exportation of wastes to a more appropriate receiving watercourse; and restricting further discharge volumes owing to surface water quality constraints.

Modelled response of the dissolved oxygen regime was most sensitive to reductions in photosynthesis and respiration factors, emphasizing the need for implementation of measures to control aquatic vegetation. Physical removal of heavy vegetation growth from the stream would improve the dissolved oxygen regime and produce a more aesthetically pleasing watercourse. To date, experimental use of aquatic plant harvesting as a water management technique in Ontario has been conducted only in lakes and has not been applied in a stream setting. Hence, this approach would have to be regarded as experimental and should be preceded by an indepth study of its feasibility.

Removal of phosphorus from the Stratford effluent to lowest possible levels during the critical growth period (May through September) would have indirect positive effects on dissolved oxygen levels by reducing the continuous supply of nutrients available for growth to minimal amounts. Municipal sources accounted for 56 percent of the annual phosphorus loading in the Avon River prior to the implementation of the phosphorus removal program in 1973. It is calculated that this figure has been reduced to 42 percent following reductions in phosphorus levels in the treated sewage effluent.

It must be noted that 80 percent of the annual contribution of phosphorus from Stratford originates from sources other than the municipal waste treatment plant, including sewage bypasses, urban runoff and minor point source inputs. However, as indicated in Table 4, the continuous year-round phosphorus input from the sewage facility assumes greater significance in the summer growth period for aquatic vegetation, as the

Stratford plant represents the major phosphorus source even with conventional phosphorus removal during this period.

TABLE 4: MONTHLY CONTRIBUTION OF PHOSPHORUS FROM THE STRATFORD SEWAGE
TREATMENT PLANT* RELATIVE TO THE OVERALL BURDEN IN THE AVON
RIVER AT ITS MOUTH

Month (1972)	Ratio-STP Load to Avon Yield at North Thames R.
January	0.14
February	**
March	0.05
April	0.05
May	0.50
June	0.96
July	1***
August	1***
September	1***
October	0.28
November	0.23
December	0.08

* Assuming 80 percent phosphorus removal at the plant.

** Figures inaccurate owing to interference in streamflow measurement by ice.

*** Yield figures low because of stream storage of phosphorus through vegetation uptake.

A positive response of dissolved oxygen levels to reductions in oxygen demanding loadings was predicted at 0.7 miles downstream from the Stratford STP. Due to the overriding effects of the extensive beds of aquatic vegetation on the dissolved oxygen regime during the study, negligible improvement occurred further downstream. Decreasing the amount of aquatic vegetation, by implementing measures described above, increases the relative influence of organic loadings on the improved oxygen regime. Recent observations on the performance of the Stratford sewage treatment plant indicate excellent reductions in the carbonaceous portion of the organic load to the lowest practical levels, although further imrpovement in the oxidation of the nitrogenous fraction is possible.

The exportation of treated water to a receiver with a greater dilution capacity would essentially eliminate flow in the stream during the summer when natural streamflow would be expected to be intermittent.

If treated sewage were redirected to the North Thames River, downstream from the proposed Glengowan dam site, a significant stress on water quality in the reservoir would be eliminated. However, such a proposal would involve a costly 10 mile long pipeline and the discharged water would bypass the Avon River and those who have come to expect a continuous streamflow, (e.g. farmers for livestock watering), would be forced to find alternate supplies.

Studies indicate that water quality objectives may never be totally attained on the Avon River downstream from Stratford. Such constituents as chloramines, phenols, and heavy metals may never be completely removed and even low levels of these components assume significance when essentially no dilution is afforded by the receiving stream during low flows. However, by adopting the first three approaches as discussed earlier, the existing situation may be improved and the degree of impairment lessened. Keeping in mind the objective of upgrading water quality where impaired and recognizing that an acceptable sewage to streamflow ratio is grossly exceeded at present sewage flows, the discharge of increased waste loadings to the Avon River should not be allowed. Studies should be conducted to determine the significance of the municipal storm drainage portion of the organic load, and steps taken to correct problems where they exist.

Tavistock

The water management problems in the Thames River upstream from London are typified by the problems encountered downstream from Tavistock as follows.

From Zorra Swamp to the Pittock reservoir, the problem of water quality impairment appears to be related to a combination of low streamflows and rural runoff. Water management options to upgrade water quality in this reach include flow augmentation and the adoption of alternatives related to rural areas.

Water resource management options directed towards the solution of problems related to rural areas are particularly important in this reach. One reason is the intensity of cattle and swine production in East Zorra Township, which occupies the major drainage area of this stretch

of the river. Federal livestock census figures for 1971 reveal that East Zorra Township carries the largest numbers of cattle and hogs of any township in the Thames River basin. Hence, options described in greater detail previously in this chapter, such as the restriction of range cattle access to watercourses and close surveillance of manure handling practices and water quality of municipal drains, are of particular relevance to this area.

Considering the low waste assimilative capacity of the stream and conflicts with downstream uses, serious consideration should be given to limiting urban development in this headwater reach. Constraints on waste discharge methods at Tavistock are imposed mainly by critically low streamflows from May through October and by downstream water uses. During these six months, monthly average streamflows may drop below 1 cfs and discharged wastes would become partially entrapped in the Gordon Pittock reservoir. By the year 2001, it is projected that the population of Tavistock will have increased to approximately 2,100. At the present population (1,400) and operating under the constraints mentioned previously, water quality objectives are barely achieved in this reach of the Thames River. In addition, a significant portion of the phosphorus load from the village is discharged at the same time that water is being stored in Gordon Pittock reservoir. Population expansion would therefore require greater waste storage capacity, a change in discharge timing and/or improved treatment to reduce oxygen consumption and phosphorus loadings to acceptable levels.

The option of total retention-spray irrigation, based on the 2,100 population figure, would involve the acquisition of an estimated 200 acres for spray purposes, 55 acres for 265 days of storage plus additional acreage for berms and a buffer zone.

One alternative course of action would be increased retention for discharge at varying rates during March and April, which would involve lagoon expansion from 14 to approximately 60 acres. Although water quality objectives for the Thames River from Tavistock to Woodstock would be attained by this approach, the total load of bacteria and nutrients would be discharged at a time when the Pittock reservoir receives and stores the spring freshet, thus causing possible further interference with aesthetic and recreational uses of the reservoir.

Storing treated wastes from May through October and discharging daily flows from a secondary treatment plant from November through April, only when streamflows exceed 2.2 cfs, is the most practical approach to future waste treatment at Tavistock. Wastes stored in a 35-acre lagoon would be discharged as streamflows permit. The annual burden of nutrients and bacteria originating from Tavistock and being taken into storage by the Gordon Pittock reservoir would be significantly reduced and water quality objectives in this reach of the Thames River would be met.

Conventional phosphorus removal incorporated into the latter two options would reduce Tavistock's present phosphorus contribution to the river from a measured 23 percent of the annual stream burden to 6.6 percent of the burden, using population figures for the year 2001. Since this source represented only 6.6 percent of the annual stream burden entering Pittock, efforts would be better expended on the control of rural sources, which account for over 90 percent of the phosphorus burden in the stream.

The three waste treatment options suggested for Tavistock are typically considered when a receiving stream is fast approaching or already beyond its ultimate capacity to accept wastes. To further tax the receiving watercourse by loadings beyond the 2001 design figure would seriously jeopardize water quality both in the immediate downstream reaches of the Thames River and in the Gordon Pittock reservoir. Since expansion of Tavistock could proceed only at the expense of well over 200 acres of farm-land or construction of two costly sewage facilities, the most logical option available to the Village of Tavistock is the application of growth restraints.

Construction of the Zorra Swamp reservoir is a possibility. If it were built for augmentation purposes, waste treatment requirements for Tavistock would not be as stringent as described above; also, it is unlikely that immediate growth restrictions would be required. The additional 10 cfs of flow provided during the critical low flow period from the reservoir would allow discharge of treated sewage in this period and possibly allow the use of conventional sewage treatment systems, or reduced effluent storage requirements. However, construction of this dam for flow augmentation alone is not felt to be justified.
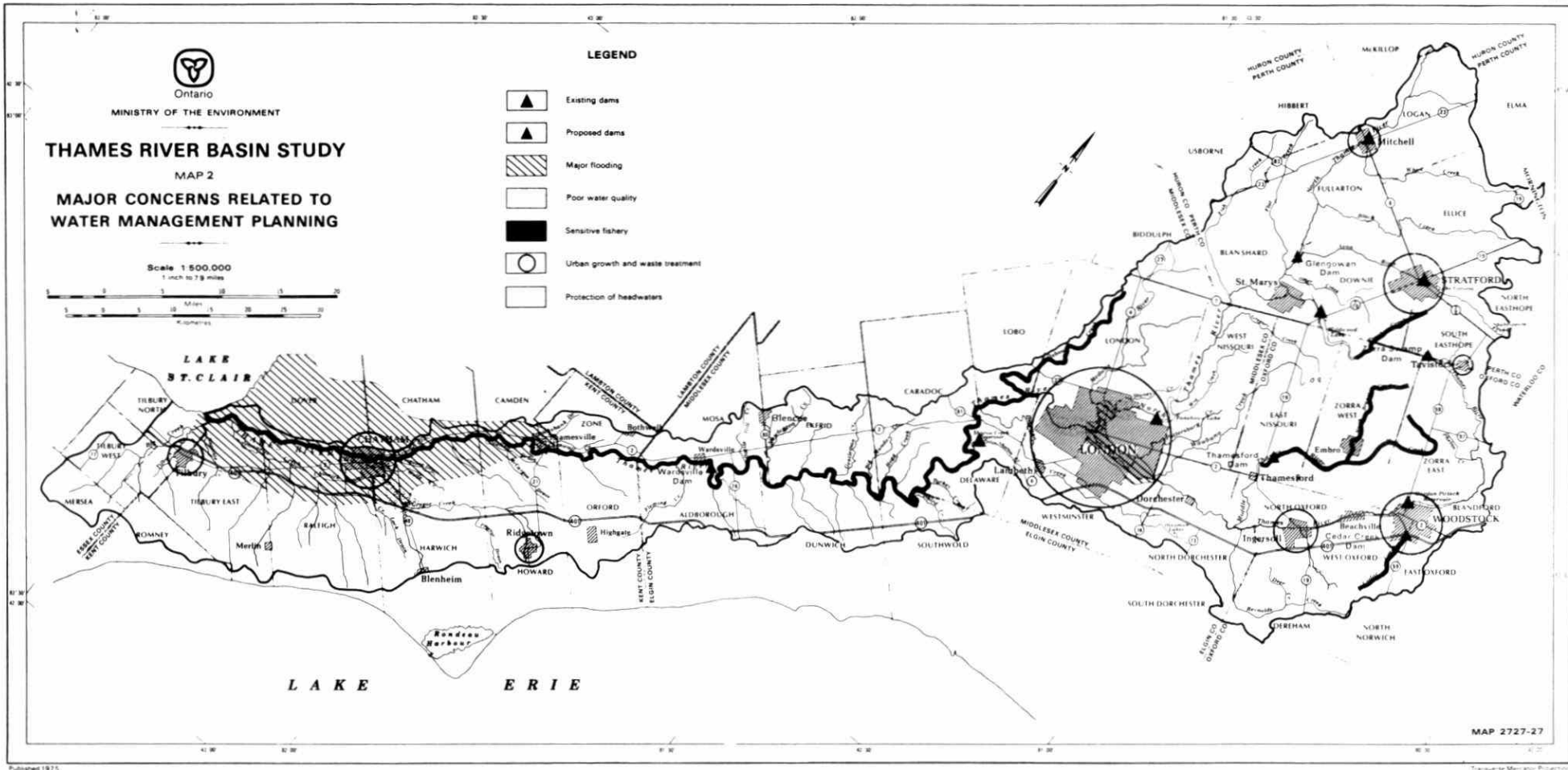
Conclusion

The preceding illustrates the reasons for requiring high quality effluents and illustrates the methodologies which have led to the definition of the generally high degrees of municipal treatment needs.

In this regard, let me make two points. The sophisticated modelling approach used for the Thames River in the London area reflects the complexities of the resource problems that existed and the fine tuning required to solve them. As far as choosing assessment techniques is concerned, their degree of sophistication should be directly related to the complexity of the resource problems at hand. To my mind, there are few situations in southern Ontario which require simulation modelling of the type used in the Thames River. Simpler steady-state modelling or even dilution models suffice in many cases and are considerably less expensive.

On the other hand, the Thames River study well illustrates that municipal wastes are but one of many sources of pollution. General urban storm and rural drainages are perfect examples of these other sources. Where adequate dilution or waste assimilative capacity exists, these sources may be relatively unimportant from a local point of view. Within the Great Lakes context, their significance is being investigated through the IJC Reference PLUARG. But where water use is intensive and competes for priority, waste control becomes more important and so does the significance of these sources. At present, we are not in a good position to control these sources. In fact, we are not even sure of the significance of urban storm drainage on water quality and the controls are therefore being put on the so-called controllable point sources, leading in many cases to requiring high quality, expensive treatment.

What is the return for our dollars spent in Ontario? In general, it has been excellent. Public health problems are almost non-existent and water quality has improved. However, we seem at a crossroad where caution should be exercised before we proceed. The intensification of water uses such as waste disposal, recreation, water supply, and fish and wildlife requires us to be more and more careful in controlling and managing our water resources. That management has to become an integral part of provincial land use planning to be ultimately effective. Our return for the dollars spent on waste treatment diminishes in those cases where inadequate

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MAP 2
MAJOR CONCERNS RELATED TO WATER MANAGEMENT PLANNING

Ontario
MINISTRY OF THE ENVIRONMENT

THAMES RIVER BASIN STUDY

MAP 2

MAJOR CONCERNS RELATED TO WATER MANAGEMENT PLANNING

Scale 1:500,000
1 inch to 7.9 miles

Miles

Kilometres

LEGEND

▲	Existing dams
▲	Proposed dams
▨	Major flooding
☐	Poor water quality
■	Sensitive fishery
◯	Urban growth and waste treatment
☐	Protection of headwaters

LAKE ST. CLAIR

LAKE ERIE

Rondeau Harbour

TILBURY NORTH, TILBURY WEST, TILBURY EAST, MERSEA, ROMNEY, RALEIGH, HARWICH, HOWARD, Blenheim, Merlin, Ridgetown, Highgate, Tilbury, CHATHAM, DOVER, CAMDEN, ZONE, Thamesville, Bothwell, ORFORD, ALDBOROUGH, DUNWICH, SOUTHWOLD, MOSA, EKFRID, Wardsville, Wardsville Dam, DELAWARE, WESTMINSTER, Blencoe, CARADOC, LOBO, LONDON, Lambeth, Dorchester, NORTH DORCHESTER, SOUTH DORCHESTER, Thamesford, Thamesford Dam, NORTH OXFORD, Ingersoll, WEST OXFORD, EAST OXFORD, DEREHAM, NORTH NORWICH, WOODSTOCK, Beachville, Cedar Creek Dam, BLANDFORD, EAST NISSOURI, WEST NISSOURI, ZORRA WEST, ZORRA EAST, Embro, St. Marys, Glengowan Dam, DOWNIE, BIDDULPH, BLANSHARD, USBORNE, FULLARTON, Mitchell, HIBBERT, McKILLOP, LOGAN, ELMA, ELLICE, SOUTH EASTHOPE, NORTH EASTHOPE, Tavistock, STRATFORD, Tavi Dam

HURON COUNTY / PERTH COUNTY
PERTH CO. / OXFORD CO.
MIDDLESEX COUNTY / OXFORD CO.
LAMBTON COUNTY / KENT COUNTY
LAMBTON COUNTY / MIDDLESEX COUNTY
ESSEX COUNTY / KENT COUNTY
KENT COUNTY / ELGIN COUNTY
MIDDLESEX COUNTY / ELGIN COUNTY
ELGIN CO. / OXFORD CO.
WATERLOO CO. / OXFORD CO.

MAP 2727-27

waste receivers are available. The point I am making is that we all too often neglect to consider the alternative of no growth, or resolving local public health problems caused by faulty septic tanks by eliminating these problem septic tanks. Instead we often build expensive sewage treatment plants which have marginal wastewater receivers and which become even more marginal if the municipality is allowed to grow.

The Thames River Water Management Plan is a guide to future development of that watershed. Provincial and municipal planning agencies welcome these guidelines on water resource limitations as important information in their work to plan land use. I quite frankly hope that this will eventually eliminate the need for high quality effluents.

DISINFECTION

by

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GENERAL CONSIDERATIONS IN WASTEWATER DISINFECTION

The objective of disinfection, in wastewater treatment, is the destruction of pathogenic microorganisms before discharge of the wastewater to a receiver.

The classical justifications for disinfection are the protection of receiving waters used for swimming or other total body contact sports, and as sources for public water supplies. Water treatment plants have their own stringent lines of defence against pathogens. For the protection of such supplies, sewage disinfection has been regarded as a supplementary line of defence. The justification for the concept of supplementary defence is that, at the least, the sewage plant effluent represents a large potential point source of discharge of pathogens. In jurisdictions where disinfection is practised, it is presumed that the process is effective in eliminating pathogens and also that the absence of disinfection may lead to massive outbreaks of waterborne disease, due to pathogens conveyed to the receiver in the sewage plant effluent.

Table 1 lists some diseases which are capable of transmission by water. The list is formidable and includes diseases caused by bacteria, parasites and viruses.

Complete sterilization of the wastewater is not the objective of disinfection. Even if this were technically feasible, it would not be desirable. Ideally, disinfection would inactivate pathogens without the destruction of beneficial organisms. Such selectivity is never likely to be achieved. Consequently, beneficial and harmless organisms will inevitably be eliminated during the disinfection process, and useful agents will be drawn from those with 'wide-spectrum' disinfecting abilities.

TABLE 1. SOME DISEASES KNOWN TO BE TRANSMITTED BY WATER (1)

Typhoid Fever

Cholera

Amoebic Dysentery

Gastroenteritis

Schistosomiasis

Paralytic Poliomyelitis

Aseptic Meningitis

Respiratory Diseases

Enteritis

Infectious Hepatitis

Worms - Roundworms, Hookworms

Disinfection, as conventionally applied in wastewater treatment, is a 'tail-end' unit operation. Merely because the process is at the end of the flow sheet does not mean it should be regarded as the last chance for making physical-chemical changes to the wastewater. Ideally, the wastewater would already have received treatment to such a degree that disinfection would not cause sinificant changes in effluent quality, upgrading effects included.

Upgrading denotes a decrease in parameters such as BOD_5, TOC, TSS, colour, turbidity. It is usually the case that the upgrading process competes for the disinfecting agent to the detriment of the disinfecting function. This increases the total application rate required for a given level of disinfection, which increases the cost. Additionally, the upgrading effects usually make the process more difficult to control, since there is no certainty as to what purpose the disinfectant served whilst being consumed.

The ideal of little interaction between disinfectant and wastewater is difficult to meet. Powerful oxidizing chemicals, such as chlorine and ozone, not surprisingly, do cause such change reacting with both inorganic and organic constituents in the wastewater. By contrast, one advantage of gamma irradiation, for wastewater disinfection, is that it does not cause significant change at the dosage required to be effective. This seems to be true as far as easily measurable change is concerned, but more subtle effects can take a long time to become apparent.

What are the requirements of an ideal disinfectant? Table 2 lists some of them, and hopefully includes all the major ones. Doubtless, other requirements could be added, but the list is already long, and no single disinfection process available for use in wastewater treatment today meets them all.

TABLE 2. REQUIREMENTS FOR AN IDEAL DISINFECTION AGENT

Wide Spectrum Disinfecting Action	Bactericidal Sporicidal Cysticidal Viricidal
Economic	Readily available Relatively inexpensive
Practical	Easy to apply Dosage can be readily measured Process can be readily controlled Results can be readily predicted Reaction products - non toxic, biodegradable, non mutagenic

For Chemical Disinfectants	For Radiation
Chemically weak Soluble in Wastewater Chemistry in Wastewater is known	Chemically weak Adequate penetrating Power in Wastewater

Source: Classification based on Reference (2)

What help can the preceding treatment stages give to the disinfection process? Irrespective of the disinfection process used, the most important single aid that the preceding treatment can offer is a population of pathogens already drastically reduced. A benefit that can be expected to accompany such treatment is that the effluent will be as free as possible of substances which will interact chemically with the disinfectant. The inclusion of chemical coagulation and filtration steps within the treatment process will generally assist both in minimizing the required dosage of disinfectant, and in reducing the number

of pathogens present before disinfection. An extreme example of the possibilities of treatment for inactivation of viruses is shown in the data presented in Table 3.

TABLE 3. VIRUS SAMPLING 1969, AT SOUTH TAHOE PUBLIC UTILITY
DISTRICT WATER RECLAMATION PLANT (3) VIRUSES
RECOVERED (PLAQUE FORMING UNITS)

Date	Primary Effluent	Secondary Effluent	Carbon Column Unchlorinated Effluent	Chlorinated Final Effluent
May 29	3	0	1	0
June 5			0	0
June 12			0	0
Aug. 20	3	18	*	0
Aug. 27			*	0
Sept. 11			0	0
Sept. 18	179	14	9	0
Sept. 25	*	430	0	0
Oct. 2	207	320	0	0

* No reliable data

Virus examination carried out by EPA Laboratories - Cincinnati

These data are taken from a recently reported study (3) relating to the Lake Tahoe Advanced Waste Treatment plant. This plant includes conventional activated sludge treatment, high-lime chemical clarification, experimental ammonia stripping to partial plant capacity, recarbonation, mixed media filtration, granular carbon adsorption and chlorination. Note the large decrease in viruses recovered between the secondary effluent and the carbon column effluent. This plant also achieves excellent bacterial disinfection with a chlorine dose of 2-3 mg/l despite the presence of 2-15 mg/l of ammonia. This is not surprising, considering the total coliform count (MPN) in the undisinfected wastewater is usually below

3000/100 ml. This low value compares with a typical value of 100,000/100 ml from a conventional activated sludge plant. Disinfection of Lake Tahoe wastewater obviously presents few problems, compared with an activated sludge effluent.

MEASURING THE EFFICIENCY AND END RESULT OF DISINFECTION

Having decided that the elimination of pathogens is the chief objective of disinfection, it is necessary to select criteria for assessing the efficiency and end result of the disinfection process actually used.

At this point, significant problems appear and obstruct a seemingly simple desire. Firstly, there is the matter of selecting a suitable organism or organisms on which to base the assessment. There is no shortage of candidate organisms, but some are known to be present only infrequently whilst others require highly specific concentration, isolation and enumeration techniques.

Above all, there is the wide spectrum of resistance shown to a particular disinfectant by various organisms and the knowledge that the order of resistance may vary for a specific group when comparing different disinfectants. This suggests that more than one organism must be used in assessment of efficiency, and at the level of basic research, action of potential disinfectants is invariably tested against a wide variety of organisms.

At the practical level, in the WPCP plant, the traditional indicator organisms are used, usually the total coliform and/or fecal coilform groups. Whilst these give an indication of the fate of pathogenic vegetative bacteria, they provide no guidance as to the fate of the generally more resistant organisms, such as spores and viruses. It is not the object of disinfection merely to kill coliforms, and there is an urgent need to establish some consensus organism to measure the effects of disinfection upon the numbers of more resistant organisms, particularly viruses.

For the present, attempts to regulate disinfection efficiency by bacteriological criteria relate principally to total coliforms and fecal coliforms. For chlorinated wastewater, at least, enumeration of fecal coliforms can give widely different results depending on the methodology

employed - Most Probable Number or Membrane Filter. As an example, consider the data presented in Figure 1. This is a graphical comparison of fecal coliform counts on the same chlorinated wastewater as determined by MPN and membrane filter techniques. The methodology employed is conventional and is described fully in the original reference (4).

On the whole, the MPN counts are greater than those for the membrane filter. In practical recognition of these problems, the Environmental Protection Agency in the United States has recently proposed that MPN procedures be mandatory for fecal coliform determination on chlorinated wastewater samples, when assessing compliance with Federal standards for disinfection (5).

Enumeration of total coliforms does not seem to be subject to the same problems, and for routine assessment of the consistency of the disinfection process within a WPCP this group seems to be the most suitable.

As to what level of total or fecal coliforms is desirable in a disinfected effluent, there is a wide range of bacteriological standards in existence. The EPA has instituted a standard of 200 fecal coliforms/ 100 ml as a monthly average. California has very strict standards that require the median count of total coliforms to be 23/100 ml where discharge takes place to confined waters.

This type of standard attempts to relate the presence or absence of a particular number of coliforms to a definable health hazard in either the treatment plant effluent or the receiver. In either case, there may be an allowance for dilution effects. The origins of these standards vary in different jurisdictions. Some are based on legal precedents, others are based on statistical studies attempting to relate the likely presence or absence of pathogens to a particular number of coliforms. In any event, as mentioned previously, the number of coliforms in a wastewater, taken as an isolated piece of information, is inadequate as a measure of the true efficiency of disinfection.

Irrespective of which organism or group of organisms are chosen for assessment of the disinfection process, no bacteriological counts will be available in time to provide input for process control. Nor can a truly continuous record of results be readily obtained. This is a serious disadvantage, since inferential methods of control must be sought.

FIGURE 1. COMPARISON OF MEMBRANE FILTER AND MPN
TECHNIQUES FOR RECOVERY OF FECAL COLIFORMS
FROM CHLORINATED EFFLUENT
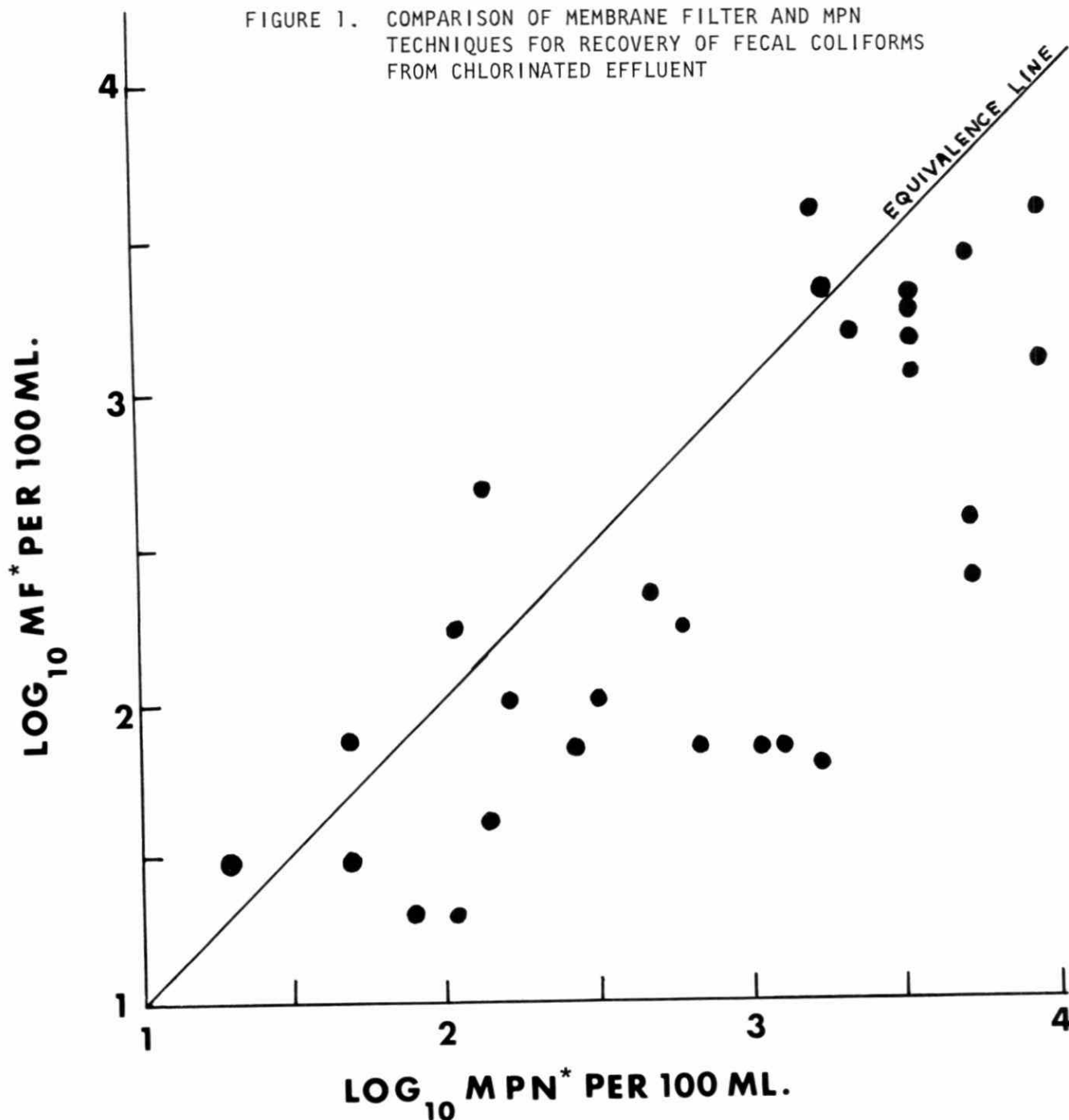
* Fecal coliforms determined by membrane filter technique, using m-FC
broth (BBL) with 0.8 percent agar (Oxoid) added.
Incubation accomplished by sealing in plastic bags and immersion in
a water bath at $44.5^{\circ}C$.
Most probable number for fecal coliforms determined by inoculation
of positive lactose tubes, from the presumptive test, into EC broth
and incubation at $44.5^{\circ}C$ for 24 hrs.

For these reasons, in jurisdictions where chlorination is mandatory, a minimum disinfectant residual plus contact time is often used as the primary evidence of proof of disinfection. A continuous record of residual can be obtained, when required, and if the minimum contact time and residual are known, there is some assurance of disinfection having occurred.

Certainly, no alarm bells ring as pathogens leave the WPCP. Along with a more resistant consensus organism than the coliform group, there is a real need for some test, method or technique that would, at the least, speedily indicate changes in "total plate count".

The above considerations are intended to highlight the fact that disinfection is a complex and inexact task in wastewater, and not an exact science. It also follows that new disinfection processes and agents must be thoroughly evaluated on as many organisms as possible, and of necessity, at present, the results correlated against the indicator groups.

Notwithstanding all of these reservations, well controlled chlorination can eliminate in the range of 99%-99.9% total coliforms, to leave about 1000/100 ml in the effluent, and counts possibly as low as one-tenth of that number as fecal coliforms, even with moderate chlorine residuals and retention times.

Experience in California (6) indicates that in warm weather climates at least, much lower counts (less than 23/100 ml total coliforms) can be consistently achieved at high residual levels (3-5 mg/l) and longer contact time (30-45 min). High chlorine residuals create their own toxicity problems and one consequence in California has been increasing use of dechlorination techniques (7).

PRESENT PRACTICE - CHLORINATION

Chemistry of Chlorination in Wastewater

This is a complex subject and only the bare essentials will be addressed, as relating to wastewater from conventional secondary treatment containing 10-30 mg/l of ammonia.

Chlorine gas is soluble in water (7160 mg/l at $20^{\circ}C$ and 1 atm) and hydrolyzes rapidly to form hypochlorous acid (HOCl) (Table 4, Equation 1).

When the chlorine dissolves in water there is an accompanying lowering of solution pH. If the pH of the solution is adjusted to 5.0 or below all the HOCl is in the undissociated form. This is the most powerful disinfecting species and in this condition the chlorine may truly be described as a wide spectrum disinfectant (8). Between pH 5.0 and pH 8.5, there is a sharp change from undissociated to dissociated hypochlorous acid. At pH 7.5 or above, hypochlorite ions (OCl^-) predominate. Hypochlorite is a relatively poor disinfectant compared with free HOCl.

TABLE 4. FUNDAMENTAL REACTION OF CHLORINE IN WASTEWATER (1)

Equation				Species
1	$Cl_2 + H_2O$	$\rightarrow$	$HOCl + H^+ + Cl^-$	Hypochlorous Acid
2	$HOCl + NH_3$	$\rightarrow$	$NH_2Cl + H_2O$	Mono-Chloramine
3	$HOCl + NH_2Cl$	$\rightarrow$	$NHCl_2 + H_2O$	Di-Chloramine
4	$HOCl + NHCl_2$	$\rightarrow$	$NCl_3 + H_2O$	Tri-Chloramine

In typical wastewater disinfection chlorine gas is dissolved in a small fraction of the total volume of wastewater to make a concentrated solution up to 3500 mg/l in strength, at low pH. This solution contains HOCl as a predominant species.

The concentrated solution is then mixed with the bulk of the wastewater. At this point the reactions of chlorine with ammonia in the wastewater become very important. The possible reactions are given in Table 4, Equations 2, 3 and 4. The end products are the chloramines which have disinfecting properties much weaker than those of hypochlorous acid.

The dominant species formed depends upon the pH and ammonia: chlorine ratio. At pH 7-8 and $17^{\circ}C$-$20^{\circ}C$ for chlorine:ammonia weight ratios of 5:1 or less, essentially all the free chlorine will be converted to mono-chloramine. Under these conditions, the reaction can go to completion in less than 1 second given efficient enough mixing (1).

Mono-chloramine is a much less potent disinfectant than hypochlorous acid. It has been shown (9) that in the pH range typical for wastewater, about twenty-five times the amount of combined chlorine is

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required relative to hypochlorous acid for the same germicidal efficiency. Additionally, mono-chloramine is a much poorer viricide (8).

It has been contended by some investigators that because of the rapidity of the reaction to form chloramines, the free chlorine is present in the wastewater too briefly to play a significant overall part in the disinfection process. Other investigators have argued the contrary - namely that a significant contribution towards elimination of more resistant organisms occurs during this very period. Regardless of what happens in the first few seconds after mixing, it is certainly the case that for practically the whole of the contact chamber detention period there is no free chlorine present.

During the first period after mixing of the concentrate, the bulk of the wastewater "chlorine demand" is being met. By "chlorine demand" we are referring to "side-reactions" of the chlorine with both inorganic and organic matter, living, dead and inanimate.

The reactions with inorganic reducing agents such as Fe^{++}, Mn^{++} and H_2S occur quickly in stoichiometric proportions. The reactions with organic compounds are generally slower, and the extent to which they proceed depends on the amount of excess chlorine. The total demand is a function of concentration of reactants, temperature and time. That portion of the chlorine used for bacterial kill forms part of the variable demand. Reactions occurring with organic nitrogen are a disguised part of the variable demand. The end products of reactions with organic nitrogen report as chloramine when the amperometric method is used to determine combined chlorine species. N-chloro compounds thus formed in wastewater chlorination generally have little or no disinfecting value, but the amperometric technique cannot distinguish them from any true di-chloramine which may be formed by the reactions of chlorine and ammonia.

Temperature Effects

Wastewater pH, and ammonia content are not subject to ready alteration in operating plants. Temperature variations are also inherent in treatment plant operation in Canada. Lowering the temperature of disinfection can seriously impair the efficiency of the process. In one series of laboratory tests (10) with mono-chloramine using E. coli and

S. typhosa as test organisms, it was shown that following a decrease in temperature from $20^{\circ}C$-$25^{\circ}C$ to $2^{\circ}C$-$6^{\circ}C$ it required 9 times the exposure or 2.5 times the quantity of chloramine residual to produce a 100% kill.

Effect of Contact Time and Residual

These two parameters are recognized as being interchangeable to some extent to obtain the same coliform kill. There have been many attempts to model the disinfection process and predict the end result.

The recent model proposed by Collins, Selleck and White (11), based on results obtained in a stirred batch reactor, assigns equal importance to residual and contact time.

The equation is:

$$Y/Yo = (1 + 0.23 \; ct)^{-3}$$

where: Y = MPN in the chlorinated wastewater at the end of time t.

Yo = MPN in the wastewater prior to chlorination.

C = total chlorine residual, mg/l.

t = contact time, minutes.

The equation obviously requires extensive field evaluation in continuous systems before being considered seriously as a design tool. However, most models proposed for chlorination have included only coliform counts, chlorine residual and contact time as variables, within the disinfection process.

Fifteen to forty-five minutes seems to be the range of contact times in common use for a defined peak flow for secondary treatment plant effluents.

DESIGN FOR AN OPTIMUM SYSTEM

Initial Mixing

It has been shown by numerous investigators (11, 12) that thorough mixing of concentrated chlorine solution and the bulk wastewater produces greatly enhanced disinfection effects. Various explanations have been suggested. One suggestion is that significant enhancement of disinfec-

tion due to HOCl occurs during the period of reaction of chlorine and
ammonia to chloramines, especially when extreme turbulence or shear
is occurring due to intense mixing. A second theory is that intense
mixing and shear breaks up clumps of microorganisms enabling more effective
disinfection.

A third theory which has been advanced is that the organic
chloramines which are formed have the effect of "poisoning" residuals of
mono-chloramine even where the residual is only a few minutes old. From
this it follows that good mixing is needed immediately to maximize the
use of fresh chlorine residual.

Whatever the explanation, good initial mixing is necessary for
optimum disinfection. Some mixing devices which have been suggested or
used are listed in Table 5.

TABLE 5. MIXING DEVICES FOR CHLORINATION

Pressure Conduit in turbulent flow

Hydraulic Jump - in a flume, channel or at a submerged weir

Pipeline Mixer

Mechanical Turbo Mixer - in a separate mixing chamber

Venturi - with chlorine applied at the throat

Eductor - with chlorine solution as the induced fluid

Contact Chamber Design

A major factor in best practice is knowing the detention time
in the contact chamber with some certainty. This suggests a chamber with
minimum "dead" areas of no flow, so that nominal and actual detention
times are very close. Additionally, the chlorine residual "poisoning"
theory suggests that back-mixing should be avoided as far as possible not
just because of wasted volume, but because back-mixing is positively
harmful to fresh chlorine residuals. These considerations suggest a
contact chamber with highly developed plug flow characteristics.

Several recent studies (6, 11, 13) have investigated different
types and configurations of contact chamber. The closed pressurized

conduit and the longitudinally baffled contact chamber seem to be the easiest to design well. A length:width ratio of 40:1 or greater (13) has been recommended to develop full plug flow characteristics. Careful attention to hydraulic factors such as entry and exit locations and geometry is also essential.

Finally, the design of the contact chamber should allow for easy flushing and cleaning for solids removal on a regular basis.

Process Control

The technology of process control for chlorination relates to maintaining a preset total combined chlorine residual at the end of the chlorine contact chamber detention period. Once the WPCP is built, and in service, chlorine residual is the only factor that can be readily adjusted.

At the crudest level of control, the residual can be maintained by manually setting the rate of chlorine addition. For good control using this method the implication would be that neither hydraulic flow nor chlorine demand vary significantly. Since flow alone can easily vary over a 4:1 range, it is obvious that either gross overdosing must be practised some of the time or residual will fall below the minimum acceptable value some of the time.

Next in complexity, there is flow-paced chlorine application in which chlorine is added in a definite ratio of hydraulic flow. This offers the possibility of much better control but is still far from perfect since the residual maintained is usually a small proportion of the total chlorine applied. For a chlorine residual of 0.5 mg/l and application rate of 5 mg/l the residual is only 10% of total application rate. If the control is set up based on peak chlorine demand, some overdosing will occur at times, but with good surveillance it is possible to maintain a minimum residual more consistently than with manual control.

The most complex, and theoretically the best residual control system, is compound loop control. In this system one or more amperometric residual analyzer controllers, plus a flow measuring device, are used to control the end-of-contact-chamber residual.

Such a system is sophisticated and expensive, and only suitable for large plants. Compound loop control has found greatest use in waste-

water treatment plants on the West Coast of the United States where chlorine residuals in excess of 4 mg/l are not uncommon (6). In such applications, in a well designed system it is possible that the systems can justify their cost in terms of chlorine savings.

The method of analysis used for chlorine residual is obviously of prime importance in good control. The amperometric titrator using Phenylarsine oxide as a reducing agent is the best available method for residuals in the range which is common in WPCP's and should be used where-ever possible.

Effectiveness of disinfection with Chlorine

When discussing this topic in relation to conventional wastewater disinfection, it cannot be emphasized too strongly that disinfection occurs essentially by the action of mono-chloramine with only the fleeting presence of hypochlorous acid.

With respect to the disinfecting ability of combined chlorine, the position is reasonably clear. Combined chlorine is deficient as a viricide and cysticide (8, 12). It is effective on the less resistant pathogens including many organisms capable of causing serious disease in man (10), when disinfection is properly practiced, as judged by high percentage kills of the total and fecal coliform groups.

The disinfecting ability of chlorine is used at some overall cost to the receiving water. Most notably, chloramine residuals toxic to fish and other aquatic life remain, and it is this, more than any other factor that has spurred research on replacements for chlorine in recent years.

DECHLORINATION

In view of the known toxicity to aquatic life of low levels of chlorine residual, additions of a dechlorination stage may be considered desirable in some WPCP's.

Dechlorination with sulphur dioxide is the best developed and least expensive technique at the present time. It has been demonstrated on a large scale (2, 14). The chemical reactions occurring are shown in Table 6. A dilute solution of sulphurous acid is formed by dissolving the sulphur dioxide in water using a modified chlorinator as the gas

metering device. This solution is then rapidly mixed with the wastewater after the chlorine contact period is over. The reaction with the chlorine is virtually instantaneous given good mixing conditions. All of the combined chlorine can be reacted to chloride at stoichiometric addition rates of sulphur dioxide, approximately 0.9 pounds of sulphur dioxide per pound of chlorine.

TABLE 6. CHEMICAL REACTIONS TAKING PLACE
IN DECHLORINATION WITH SULPHUR DIOXIDE

With free chlorine $\quad$ $SO_2 + HOCl + H_2O \rightarrow SO_4^= + Cl^- + 3H^+$

With combined chlorine $\quad$ $SO_2 + NH_2Cl + 2H_2O \rightarrow SO_4^= + Cl^- + 2H^+ + NH_4^+$

$2SO_2 + NHCl_2 + 4H_2O \rightarrow 2SO_4^= + 2Cl^- + 5H^+ + NH_4^+$

With dissolved oxygen $\quad$ $2SO_2 + 2H_2O + O_2 \rightarrow 2H_2SO_4$

The addition of sulphur compounds to the water increases the total dissolved solids slightly and can be considered to be responsible for regenerating the ammonium ion associated with the combined residual. These disadvantages are of little consequence, provided the chlorine being destroyed is small to begin with.

Two rather more practical problems are that excess sulphur dioxide will result in an additional oxygen demand of the effluent, and that lowering of wastewater pH can occur.

Both problems can be largely avoided by close control of the addition rate of sulphur dioxide. Where effluents with high dissolved oxygen are required, a post-aeration stage can be added.

To achieve close control of a dechlorination stage necessitates automatically controlled chlorine residuals to minimize the sulphur dioxide added. Flow proportional control of the sulphur dioxide addition rate is also required.

It is obvious that such a closely controlled process is best suited to larger WPCP's.

The chief objection to the whole process of dechlorination and reaeration is the increase in complexity of the disinfection process, without any corresponding benefits in disinfection itself. The added.

costs of the whole procedure of dechlorination and reaeration can be considerable. One recent estimate (15) is that the total added cost is in the range of 1.2 times the cost of chlorination. The dechlorination step itself represents about one third of the added cost.

THE FUTURE

General

When discussing the future in relation to disinfection, there are two distinct types of requirements to consider.

The first is for an upgrading or change in the disinfection method of choice for wastewaters essentially the same in quality as those of today, with high levels of ammonia, suspended solids and turbidity, and significant "demand" for powerful oxidants. The second requirement is for disinfectants for advanced wastewater treatment. The latter types of wastewater can be very diverse in quality, and any projection of needs are guesses at best. However, the future of chlorination for disinfection in such processes is probably linked with the elimination of most of the ammonia before disinfection, by nitrification or other means, permitting chlorination to a free residual, and full use of the superior disinfecting ability of HOCl.

Most of the alternatives to chlorine for disinfection which are being investigated are potentially useful for both the requirements outlined above. Their true advantages and disadvantages will only become apparent after long and thorough testing. Some of the alternatives which have been, or are being, seriously pursued today for wastewater disinfection are shown in Table 7. Of these, gama irradiation is probably one of the more versatile. The application and control of this disinfecting agent is radically different to that of chlorine and other strongly oxidizing disinfectants and it will therefore be considered in this paper.

TABLE 7. SOME POSSIBLE ALTERNATIVES TO CHLORINE FOR DISINFECTION

Chlorine Dioxide
Ozone
Bromine Chloride
Gamma Irradiation
Ultra-Violet Radiation

Gamma Irradiation

Sterilization by gamma irradiation has been used extensively upon drugs, food and hospital supplies. Radiation at the level necessary for sterilization does not induce radioactivity and has the capability of penetration into organic particles. In laboratory studies its effectiveness has been shown to be little changed in the presence of organic matter at the concentrations found in wastewaters (16, 17, 18). These studies using radioactive Cobalt 60 showed that the technique could potentially be used for wastewater disinfection. They also established that gamma irradiation is effective in wastewater against a wide variety of organisms, including total and fecal coliforms, fecal streptococci, E. coli bacteriophage and spores. More recently, pilot plant studies (19, 20) have shown that performance on wastewater is predictable, controllable and effective.

The data referred to here are reproduced from studies conducted over an 18 month period at the Burlington Skyway WPCP, an extended aeration plant. The studies were carried out using a pilot scale irradiation system constructed by Geodel Systems Inc. of Georgetown, Ont. The Ontario Ministry of the Environment and consultants retained by Geodel carried out an evaluation of the system, which was designed for flow rates in the range of 2500-25000 I gal/day.

The irradiator consisted of a closed unit incorporating a series of annuli for flow of the wastewater with over and under baffles. Pencils of Cobalt 60, the irradiation source, were located in holders within the unit around an inner annulus. Clarified secondary effluent was pumped from the overflow channel of the secondary clarifier to the irradiator, and irradiated effluent was wasted to the plant outfall line. Further details of irradiator design, safety and fail-safe precautions are available in the original reports on the evaluation (20, 21).

For purposes of comparison with chlorine, a small-scale pilot system was set up in the WPCP with a well-baffled contact chamber, using hypochlorite solution as the chlorine source. Both the small-scale chlorine contact chamber and the gamma irradiator were operated at fixed flow rates. The flow rate to the chlorine contact chamber was arranged to give a 30 minute detention period and a residual of 0.5 mg/l (as measured by O.T.A. method) at the end of contact.

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The flow rate through the irradiator was periodically adjusted to give a different total radiation dose. Previous studies had established that disinfection effects should be proportional only to the total dose (16, 17, 18).

It was found that dosages of 100,000 rads were necessary to achieve total coliform kills comparable with those of the pilot scale chlorine system. This dosage could only be calculated from a knowledge of detention time and the quantity of Cobalt 60. It was not possible to obtain practical measurement of the level used.

A comparison of overall results at the 100,000 rads dosage level is given in Table 8. The performance of the irradiation system was superior for kill of total coliforms, slightly inferior for fecal coliforms and much less efficient for fecal streptococci. Bacteriophage was present in both of the disinfected effluents. Salmonellae were isolated only once in 12 random observations during the test period at 100,000 rads.

The chlorine system performed well during the test period showing what can be done with a well controlled, well designed, flow equalized system.

Table 9 shows the percentage kills obtained with gamma irradiation on samples from two successive weeks and the variation in indicator organism counts before disinfection, during the same period.

Determination of a wide variety of parameters during the period of operation at the 100,000 rad dosage level detected only one change in effluent quality - an increase in nitrite concentration after irradiation.

Irradiator performance with respect to percentage kill of total coliforms was also very close to previously predicted values, based on laboratory scale work (21), at various radiation dosage levels.

The ability to select the lethal dose for a particular group of organisms, based on laboratory scale experiments, and then consistently achieve that result in a continuous system, cannot be matched at present by other disinfecting agents. This consistency of results, combined with the need for only minimal operator attention, and absence of gross interaction with the wastewater, is impressive.

TABLE 8. AVERAGE PERCENTAGE INACTIVATION OF MICROORGANISMS
BY GAMMA IRRADIATION AND CHLORINE (20)

Constant flow rate to each system

Results based on a minimum of 12 random observations

	Irradiation at 100,000 Rads [5]	Chlorine 0.5 mg/l residual 30 min. contact time
Total Coliform [1]	99.31	98.22
Fecal Coliform [1]	98.05	99.77
Fecal Streptococci [1]	84.42	98.80
Total Bacteria [2]	96.96	98.31
Bacteriophage [3]	Present	Present
Salmonellae [4]	Present [6]	Present

1   Bacteriological testing based on grab samples and standard membrane
    filter techniques.

2   Bacteriological testing based on grab samples and pour plate technique.
    Samples diluted and plated onto tryptone-glucose extract agar (Difco).
    Plates incubated at 35°C for 48 hours.

3   Bacteriophage based on grab samples and MPN technique of Kott and
    Vajdic. In addition 50 ml of effluent sample was added to 50 ml
    double strength lauryl tryptose broth, followed by addition of host
    bacterium E.coli b. After incubation and plating-out of MPN method,
    presence or absence of bacteriophage in the 50 ml samples could
    be determined.

4   Bacteriological testing based on Moore swab sampling. Salmonellae
    determined by method of Yoshe-Purer et al, Water Res. 4, #3, 113-120,
    1971.

5   One rad is an energy unit defined as the adsorption of 100 ergs per
    gram material.

6   Salmonellae detected only once in 12 random observations.

TABLE 9. GAMA IRRADIATION: TYPICAL RESULTS
FOR INACTIVATION OF INDICATOR ORGANISMS

| Date | Total Coliforms | | Fecal Coliforms | | Fecal Streptococci | |
	Count/100 ml Influent	% Inactivation	Count/100 ml Influent	% Inactivation	Count/100 ml Influent	% Inactivation
Jan. 8	2.1×10^4	99.33	1.35×10^3	97.04	6.0×10^2	76.67
Jan. 9	2.5×10^4	99.04	1.70×10^3	99.41	1.02×10^2	81.37
Jan. 10	3.7×10^4	99.13	5.0×10^3	96.40	1.65×10^3	77.58
Jan. 15	1.5×10^5	99.73	9.0×10^3	99.33	8.4×10^3	94.05

Above results based on grab samples and membrane filter techniques.

The major operating task with this type of system, would be replacing some of the Cobalt 60, which decays at a predictable rate, to maintain the required dosage rate. This could be carried out on an annual basis.

Possible toxicity problems associated with irradiated effluents would need careful evaluation before applying the process widely. At the present time, the major known problem with this type of disinfection is the very high cost. Geodel have produced cost estimates (20) of 64¢/1000 gallons based on a 99.9% kill of total coliforms to leave about 1000/100 ml in the disinfected effluent. Unless this cost can be drastically reduced, gamma irradiation is unlikely to be used.

Ozonation

There is currently considerable interest in ozonation as an alternative to chlorination for wastewater disinfection.

This interest is sustained by four major factors. Firstly, there is the well documented disinfecting power of ozone, in water containing low levels of organic matter. Secondly, based on limited available evidence, ozonated effluents are apparently non-toxic to aquatic life. Thirdly, significant upgrading effects may be expected to accompany the disinfection process. Lastly, there is the additional factor that since ozone usage on a large scale has been practiced for almost seventy years, the technology of ozone generation is well developed.

It is in the field the municipal water treatment, especially in Europe (25) that ozone has found its widest use, for purposes of taste and odour control and disinfection. Facilities on a large scale have been built and are operating today. In wastewater treatment, ozone has been extensively investigated, on a pilot plant scale, to assess the upgrading potential of ozonation as a unit operation in the flowsheet of Advanced Wastewater Treatment plants (22, 23, 24).

In the field of wastewater disinfection, ozone has never become, to the present time, a serious challenger to chlorine, mainly because of the greater cost and complexity of ozonation. With the current drive towards eliminating chlorine-induced toxicity in effluents discharging to sensitive receivers, cost and complexity no longer automatically override all other factors and ozone is now being considered again.

Effectiveness of ozone as a disinfectant

Numerous laboratory studies (26, 27, 28, 29) have been carried out on ozone-demand free waters. In these, ozone has been shown to be an effective bactericide, sporicide, cysticide and viricide at low concentrations (0.1-3 mg/l), and low contact times (5 minutes or less).

In laboratory tests upon cultures of bacteria suspended in buffered water, addition of very small amounts of organic contaminants has been shown to significantly increase the total ozone dosage required for complete inactivation (26).

In laboratory batch tests on wastewater, longer contact periods and high total ozone dosages have been shown to be necessary to produce effective bacterial inactivation (26, 32) than in demand free water. Such laboratory tests have confirmed that ozone is a powerful disinfecting agent in wastewater.

In one series of tests (32) it was shown that not only did ozone destroy coliforms and fecal streptococci in wastewater more rapidly than chlorine, but also that the action of the ozone, unlike chlorine, was not significantly temperature-dependent in the range of $0°-30°C$.

A number of recent pilot plant studies have been carried out on activated sludge plant effluent, filtered and unfiltered, and effluents from physical-chemical processes (12, 33, 34).

The dosage of ozone required to achieve good disinfection on secondary effluents, as judged by acceptable total coliform and fecal coliform reductions has been in the range of 5-15 mg/l in most of these studies. Contact times varied from less than 1 minute to 20 minutes.

A wide variety of upgrading effects have also been reported. Some widely observed effects are noted in Table 10. The magnitude of each individual upgrading effect has varied widely, in different studies, as might be expected since initial wastewater quality also varied widely. A further side-benefit of ozonation is that disinfected effluent with high dissolved oxygen levels can be obtained, if considered desirable, especially where ozone is generated from pure oxygen.

TABLE 10.   UPGRADING EFFECTS ACCOMPANYING DISINFECTION BY OZONE

Reduction in Suspended Solids

Reduction in COD

Reduction in Colour

Reduction in Turbidity

Oxidation of Nitrite

Destruction of Cyanides

Destruction of ABS and LAS Detergents

Reports from different studies have ascribed varying degrees of importance in upgrading to froth floatation effects accompanying ozonation. The froth, which is generated due to the disengagement of undissolved gases from the water after ozonation, can contain appreciable amounts of suspended solids. It can be readily skimmed and removed, thus enhancing upgrading effects.

Two recent pilot plant studies (12, 34) have concentrated on assessing the viricidal efficiency of ozone in wastewaters from the activated sludge process, with the usual ammonia levels. These studies showed that ozone was highly effective, much more effective than chlorine, for viral inactivation, even at low dosages where good bacterial kill was not obtained.

A recent critical review (10) of the use of ozone for wastewater disinfection notes that most of the studies to date have been carried out on a small scale (less than 0.15 MGD), for a few months duration at most.

The methodology used in conducting these studies and the reported results have amply demonstrated three major deficiencies in current state-of-the-art knowledge and technology for application of ozone to wastewater. These are discussed below.

Mixing and contact systems

No obvious first choice of mixing or contact systems has emerged to date, although a wide variety of mixing devices, in particular, have been evaluated.

Table 11 lists some devices which are currently in use in pilot plants or have been evaluated in them. Some of them have been shown to have definite advantages in a specific case, none are suitable for universal application.

TABLE 11. OZONE MIXING DEVICES

Static Mixer Element

Sonic Mixer

Venturi Mixer

Porous Plastic Diffuser

Multi-turbine Shaft Mixer

A wide variety of contact times has been previously noted as having been used in pilot plant studies. The minimum contact time used in ozonation pilot plants is partly governed by the number of mixer-contactor stages needed to solubilize the required total dosage of ozone, efficiently and without excessive losses.

The required contact time, mixing energy and number of contact stages can all greatly affect the economics of ozonation. The criteria to permit design of large scale complete systems for optimum performance are not yet well defined.

Process control

Pilot plant studies of disinfection by ozone, and most other disinfectants, invariably operate at fixed hydraulic flow rate, primarily to make data interpretation easier. However, with full scale disinfection processes, effective mechanisms for good process control are needed.

The ability to alter the addition rate of disinfectant to the wastewater is an essential feature of good process control, needed to match changes in either or both of hydraulic flow and effluent "demand". Changes in the application rate of disinfectant must be achievable at times significant to the disinfection process itself.

Changes to match effluent "demand" present the greatest problem in current state-of-the-art knowledge of ozonation. No concepts exist

in current ozone technology, relating to wastewater disinfection, equivalent to the role in chlorination, of chlorine demand, and end-of-contact-period residual. There is no single key parameter in either the undisinfected or disinfected effluent which can be readily measured and used for control of the process. A residual may appear and it may not, more frequently not. Its presence or absence is certainly related to the quantity of ozone-oxidizable material in the wastewater, but not necessarily to the disinfection efficiency achieved. In practical situations, even where a residual does develop it often decays very swiftly, within 10 seconds of leaving the contactor in one reported study (35).

At the present time, there is no ready means of predicting the ozone dosage needed to disinfect a wastewater, and pilot plant studies are essential in each case. The only simple means of process control available is that of consistently overdosing ozone in the hope of obtaining good disinfection.

Complexity of ozone generation and use

Ozone must be generated at site. Site generation of ozone or any other chemical to be used in wastewater treatment usually means added complexity of use when compared with those chemicals which can be brought to site and stored in concentrated bulk form until required.

Ozone may be generated from air, oxygen-enriched air, or pure oxygen. Each system may have advantages for a specific case. Regardless of what gas is fed to the ozonator, the ozonation system will be complex. As an example, the elements in an ozonation system using air are shown in Table 12. Because of the need for standby capacity, at least two equipment streams or back-up chlorination equipment, will be needed.

It is inappropriate to make detailed comment on the amount of operator attention that monitoring and control of the ozone contactor and overall system will require, since there are so many possible designs of system to choose from, and a lack of operating experience at the present time. It is sure to be greater than that for a chlorine system, even with dechlorination and reaeration facilities added.

TABLE 12. DISINFECTION BY OZONATION - ELEMENTS IN
A SYSTEM GENERATING OZONE FROM AIR

Air Preparation

Air Filter

Air Blower

Air Dryers - Refrigerant Stage

- Dessicant Stage

Ozone Generation and Application

Ozone Injector

Ozone Contactor

Controls

For Ozone Application Rate

For Safety and Malfunction

Costs of ozonation

There are few reliable data to provide a basis of comparison
between the dosage of chlorine and ozone needed for equivalent disinfection.
Equivalent disinfection should be judged by the inactivation of selected
organisms, to a specified level, after disinfection.

Such data as exist indicate that the ozone dosage will
usually be more than that for chlorine, if the criterion used is the
inactivation of total or fecal coliforms (10). A number of ongoing pilot
plant studies are being conducted in Canada and the U.S.A. comparing
ozone and chlorine. These will generate additional data on relative
dosages needed for equivalent disinfection.

When used at the same dosage (as mg/l applied to the wastewater),
the cost of ozone has been estimated in one recent study (15) as being
currently between twice and four times the cost of chlorination, for
plants in the range of 0.8 IMGD-80 IMGD. On all but the smallest size
plant costed (0.8 IMGD), the projected cost of ozonation was considerably
greater than that of chlorination, dechlorination and reaeration. Any
projected costs for large scale disinfection by ozonation must be regarded
as tentative, until a few installations have been built and have operated
for some years.

Prospects of disinfection by ozonation

In the near term there are probably two promising potential applications in wastewater disinfection.

The first is in water reclamation plants, where the upgrading potential, good bactericidal and viricidal abilities of ozone may make it the disinfectant of choice.

The second is in plants using the pure oxygen activated sludge process. In this case, large scale oxygen storage or production facilities will already exist at the treatment plant, and the incremental cost of producing oxygen for use in the ozonator will be lower than usual. Gases from the ozone contactor can probably be wasted to, and used in, the aeration tanks, thus eliminating the necessity of an oxygen-recirculation system for these gases.

Widespread use of ozone for wastewater disinfection will probably not occur until such time as significant advances take place in technology relating to its effective application.

REFERENCES

1. White, G.C., "Handbook of Chlorination", Van Nostrand Reinhold, N.Y., 744 p., 1972.

2. Kinman, Dr. Riley N., "State-of-the-Art in Wastewater Disinfection", EPA Workshop on Disinfection, Wyoming, Michigan, Oct. 1974.

3. Culp, Russel L., "Breakpoint Chlorination for Virus Inactivation", Proc. of Water Resources Symposium #7, University of Texas, Austin, Texas, April, 1974, 158-165.

4. Schiemann, D.A., M.D., "Bacterial Indicators of Disinfection", Annual Meeting - National Environmental Health Association, Minn. Minn., July, 1975.

5. Federal Register (U.S.), 40, 324539, June 9, 1975.

6. White, Clifford C., "Disinfection Practices in the San Francisco Bay Area", Jour. WPCF, 46, 89-101, 1974.

7. Collins, Harvey, F., Deaner, David G., "Sewage Chlorination Versus Toxicity - A Dilemma?", Proc. ASCE, 99, EE6, 762-771, 1973.

8. Kruse, C.W., Hsu, Y.C., Griffiths, A.C., Stringer, R., "Halogen Action in Bacteria, Viruses and Protozoa", National Speciality Conference on Disinfection, Univer. of Mass., July, 1970, pp. 113-135.

9. Fair, G.M., Mums, J.C., Chang, S.L., Weil, R., Burden, R.P., "The Behaviour of Chlorine as a Water Disinfectant", J. AWWA, 40, 1051, (1948).

10. Butterfield, C.T., Wattie, E., "Influence of pH and Temperature on the Survival of Coliforms and Enteric Pathogens in Water", Publ. Health Rep. (U.S.), 61, 157-192, 1946.

11. Collins, Harvey, F., Selleck, Robert E., White, George C., "Problems in Obtaining Adequate Sewage Disinfection", J. Sanitary Eng. Div. ASCE, 97, 549-561, 1971.

12. Longley, Karl E., Olivieri, Vincent P., Kruse, Cornelius, W., Kazuyoshi, Kawata, "Enhancement of Terminal Disinfection of a Wastewater Treatment System", Proc. of Water Resources Symposium #7, University of Texas, Austin, Texas, April 1974, 166-180.

13. Marske, D.M., Boyle, J.D., "Chlorine Contact Chamber Design - A Field Evaluation", Water & Sewage Works, 120 (1), 70-77, 1973.

14. Servizi, J.A., Martens, D.W., "Disinfection, Dechlorination and Toxicity to Fish", Water Resources and Environment Conference, Toronto, Ont., April 20-23, 1975.

15. Smith, R., Eilers, R.G., McMichael, W.F., "Cost of Alternative Processes for Wastewater Disinfection", EPA Workshhop on Disinfection, Wyoming, Mich., October, 1974.

16. Ridenour, Gerald M., Armbruster, Edward M., "Effect of High Level Gamma Radiation on Disinfection of Water and Sewage", J.AWWA, 48, 671-676, 1956.

17. Touhill, C.J., Martin, E.C., Fujihma, M.P., Olesen, D.E., Stein, J.E., McDarnell, G., "Effects of Radiation on Chicago Metropolitan Sanitary District Municipal and Industrial Wastewater", JWPCF, 41, Pt. 2, R44-R60, 1969.

18. Lowe, Harry N., Lacey, William J., Surkiewicz, Bernard F., Jagger, Robert F., "Destruction of Microorganisms in Water, Sewage and Sewage Sludge by Ionizing Radiation", J. AWWA, 48, 1363-1372, 1956.

19. Woodbridge, David D., Mann, Leland A., Garnett, William R., "Application of Gamma Radiation to Sewage Treatment", Nuclear News 13 (a), 60-64, 1970.

20. Vajdic, Ann H., "Gamma Rays vs the E.coli Monster (and other tales)", Water and Wastes Engineering, 12, #4, 29-32, 1975.

21. Murphy, Dr. K.L., "Gamma Radiation as an Effective Disinfectant", Water and Pollution Control, 112, 4, 24-28, 1974.

22. Boucher, Lowndes, et al, "Use of Ozone in the Reclamation of Water from Sewage Effluent", J. Inst. Public Health Engineer (U.K.), 67, 75, (1968).

23. Wynn, C.S., Kirk, B.S., McNabney, R., "Pilot Plant for Tertiary Treatment of Wastewater with Ozone", Tech. Ser. Report #EPA-R2-73-146, EPA Washington, D.C., (Jan., 1973).

24. Besik, F., "Reclamation of Potable Water from Domestic Sewage", Water & Pollut. Cont. (Canada), 109, 46, (April, 1971).

25. Gomella, C., "Ozone Practices in France", Jour. AWWA, 64, 1, 39, (Jan., 1972).

26. Smith, D.K., "Disinfection and Sterilization of Polluted Water with Ozone", 2nd Annual Symposium on Water Research, McMaster University, Hamilton, Feb., 1967.

27. "Sensitivity of Three Selected Bacterial Species on Ozone", J. Appl. Microbiol., 26, 391, 1973.

28. Katzenelson, E., Kletter, B., Schechter, M., Shuval, H.I., "Inactivation of Microorganisms by Ozone", ACS Annual Meeting, Dallas, Texas, 1967.

29. Majumdar, S.B., Ceckler, W.H., Sproul, O.J., "Inactivation of Poliovirus in Water by Ozonation", J. Water Poll. Cont. Fed., 45 (12), 2433, 1973.

30. Fetner, R.H., Ingols, R., "Ozone Chemistry and Technology", Advances in Chemistry Series 21, American Chemical Society, 1959, p. 370.

31. Kinman, Riley N., "Water and Wastewater Disinfection with Ozone; A Critical Review", CRC Critical Reviews in Environmental Control, 1975.

32. Kinman, Riley N., "Ozone Disinfection of Wastewater at Low Temperatures", Proc. Symp. Wastewater Treatment in Cold Climates, University of Saskatchewan, Saskatoon, March, 1974, 507.

33. Clark, R.G. Lowther, F.E., Rosen, H.M., "Disinfection of Municipal Secondary-Tertiary Effluents with Ozone, Five Recent Pilot-Plant Studies", 1st International Symposium and Exposition, Washington, D.C., Dec. 2-5, 1973.

34. Pareni, J.L., Tittlebaum, M.E., "Virus Inactivation in Secondary Wastewater Treatment Plant Effluent Using Ozone", 7th Water Resources Symposium #7, University of Texas, Austin, Texas, April, 1974.

35. Nebel, C., et al, "Ozone Disinfection of Combined Industrial and Municipal Secondary Effluents Pt. II Pilot Plant Studies", Proc. Purdue Industrial Waste Conference, Pt. II, 1056, May, 1972.

APPLICATION OF FLOW EQUALIZATION

by

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INTRODUCTION

Sewage flows in municipal sewage systems typically exhibit diurnal, daily and seasonal variations. The flow variability is attributable to, among others, non-uniformity in domestic water usage, industrial discharges, storm runoff, and the physical condition of the sewer lines. In the past, sewage treatment facilities have been commonly designed on the basis of treating the average daily flow with provision for handling a portion of the normal flow fluctuations. During periods of extreme high flow, however, flows exceeding a certain level would be allowed to bypass one or more treatment stages, resulting in the discharge of partially treated effluent. In some facilities the pollutional load of yearly wet weather overflows, for example, has been estimated to be of the same order of magnitude as the total yearly discharge from secondary treatment facilities (1). To protect the quality of the environment, current efforts in Canada and abroad are aimed at reducing the discharge of untreated or partially treated wastes while maintaining high removal efficiencies at reasonable cost.

Potential alternatives for eliminating untreated discharges include flow smoothing in sanitary sewerage systems, separate treatment for overflows and flow equalization within the waste treatment plant. The first alternative can involve the installation of flow smoothing basins at key locations within the sewerage system (2, 3). Wastewater is stored in the basins during periods of peak flow. Later it is released to provide nearly constant delivery to the waste treatment plant. Although this alternative may bring about effective flow control, and presumably higher waste treatment efficiency, in many urban situations it may not be economically attractive. Often no land is readily available at key locations, therefore, the basins must be constructed underground at high costs. Furthermore, additional pumping stations must be installed for emptying the storage basins.

Separate treatment of wet weather sewer overflows has been prac-
tised at least at one location in the U.S. (4). The combined sewer over-
flow receives 15 to 30 minute treatment in contact with stored viable sludge
in aerated contact stabilization basins. Subsequently, the wastewater
receives clarification and chlorination prior to discharge. Approximately
90 percent suspended solids and 80 percent BOD removal have been reported
for this system. To date, a full evaluation of this approach has not
been published. From the available data it appears that this technique
is not without limitations. Maintaining a viable sludge between incidents
of overflow and treatment efficiency at increased hydraulic loadings on the
clarifiers during wet weather flow can create serious operational problems
in conventional treatment plants.

Considering cost effectiveness and performance, the installa-
tion of equalization basins within the waste treatment plant appears to
be a promising solution (5 - 10). Excess flow is stored temporarily in the
basin and released during periods of low flow. Possible locations in the
treatment system for siting flow equalization basins include ahead of the
primary or secondary facilities (9, 10).

For maximum efficiency, the flow would have to be regulated to
achieve a constant flowrate over the entire year. However, long-term
storage, say from one season to another, would be highly impractical.
A more rational approach is to design equalization basins with sufficient
storage volume for the modulations of the diurnal variations in sewage flow.

To date, the potential benefits of flow equalization have not
been completely evaluated. Only a partial evaluation of a few case studies
has been reported (6). A full-scale study on diurnal flow equalization
has been published by LaGrega and Keenan (9). The authors found
significant improvement of suspended solids removal efficiency in
primary sedimentation with equalization. Furthermore, the diurnal varia-
bility of effluent water quality was significantly reduced.

For effective equalization design, an accurate forecasting of
raw sewage flow is essential. The literature contains several methods
of flow analysis, the simplest being the approximation of daily average
flows by a Gaussian distribution and the diurnal variation by a sine

function (7). An inspection of existing Ontario flow records indicates that these assumptions may be erroneous. A better approach is the time series analysis of flow in terms of Fourier polynomials and frequency power spectra (10, 11). The predicted flow profile can then be used for sizing and controlling flow to and from the basin to ensure near steady state operations. Time series analysis, however, is complex, and requires a large input of data.

With this in mind, this paper discusses the development of a methodology, based on the statistical properties of sewage flow, for the design of equalization basins; the assessment of potential benefits to treatment plant performance; and, the evaluation of the potential economic benefit of flow equalization. In this regard the daily and the diurnal flow fluctuations in eleven Ontario sewage treatment plants were assessed to establish realistic design criteria for waste flow equalization practices.

IMPACT ON TREATMENT

The installation of a side-line equalization basin prior to primary treatment was examined in detail. A schematic diagram of a wastewater treatment plant utilizing flow equalization is shown on Figure 1. A defined fraction of the raw sewage is diverted and stored temporarily in the equalization basin. The basin is aerated to prevent the development of anaerobic conditions and foul odours. When the raw sewage flow falls below a predetermined level, the stored sewage is fed at controlled rates to the primary treatment unit to maintain uniform flow and empty the equalization basin. Expected advantages of equalization include better treatment efficiencies, extension of the design life of plants, and elimination of sewage bypassing. Incorporating equalization in the design of new sewage treatment plants could result in cost saving by selecting smaller primary and secondary facilities.

Due to the more uniform hydraulic and mass loading, a positive impact is expected on all treatment stages. The actual benefit of flow equalization should be established from comparative case studies of wastewater treatment with and without equalization. At the present, little full-scale operating data are available, although a number of equalization practices are currently under evaluation or at the design stage.

SCHEMATIC DIAGRAM OF WASTE WATER TREATMENT FACILITIES WITH FLOW EQUALIZATION

FLOW DIVERSION

FLOW CONTROL

PRIMARY SEDIMENTATION TANK

AERATION TANK

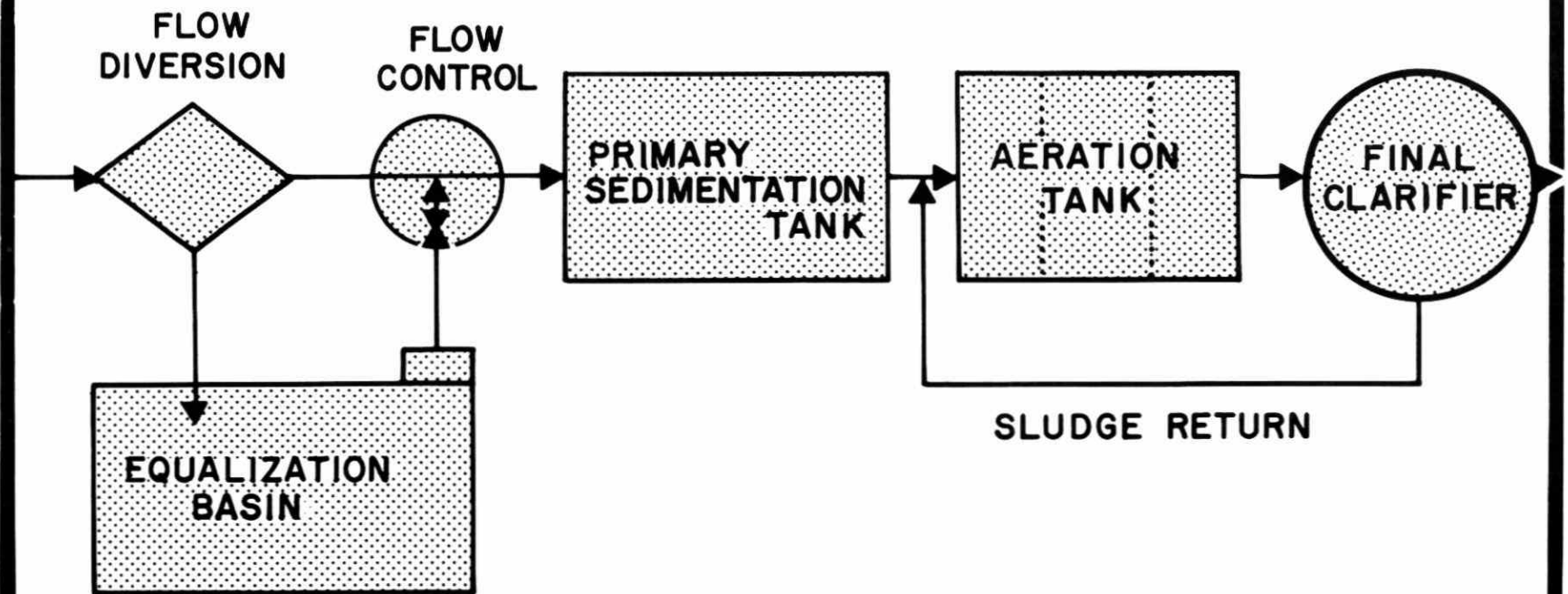FINAL CLARIFIER

EQUALIZATION BASIN

SLUDGE RETURN

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FIG. 1

The impact of flow equalization on the primary sedimentation unit was investigated by LaGrega and Keenan (9, 10). The authors performed a full-scale study at the 1.8 MGD Newark, N.Y., wastewater treatment plant. With equalization, a 47 percent removal of suspended solids was realized in primary sedimentation, compared with an average 23 percent removal efficiency without equalization. Microscopic examinations revealed that pre-aeration in the equalization basin induced preflocculation. Thus, the improved efficiency is attributable to the more uniform overflow rate in the primary, as well as the beneficial effect of induced solids flocculation.

An additional benefit may also be realized in the performance of secondary facilities. The damping of flow fluctuations will moderate shock loadings and improve the BOD removal efficiency, particularly in activated sludge units where present aeration and return sludge pumping facilities are not adequate to meet peak flow and BOD loading conditions.

Due to more uniform solids loading, flow equalization can be expected to improve secondary sedimentation by stabilizing the solids loading on the clarifiers. However, since maximum flows occur for long periods of time under flow equalization conditions, it is critical that the final clarifiers be designed for a lower 'maximum overflow rate' than for variable flow conditions (8). Comparative data concerning the effect of equalization on secondary and post-secondary facilities are very limited. One flow equalized municipal wastewater treatment plant is reported to produce a high quality effluent with BOD and suspended solids levels less than 4 mg/l and 5 mg/l respectively (6).

In a recent full-scale study (12) the effect of in-line and side-line equalization on the performance of municipal wastewater treatment plants was assessed. Both systems were highly effective in providing uniform flow to the treatment units, but the effect of equalization on both primary and secondary effluent water quality was less significant than other operational variables. The results are somewhat contradictory with previous observations (6, 10). Post-secondary filter performance, however, improved with equalization practice.

EQUALIZATION BASIN DESIGN

Determination of equalization volume requirements is often based on the characteristic diurnal flow pattern. The most common procedure (6, 8) used is illustrated on Figure 2. In this hydrograph, cumulative flow over a 24-hour period is plotted against time. The straight line represents equalized flow conditions. At any particular time the instantaneous flowrate is given by the slope of the curves. From time "o" to "a" the uncontrolled flowrate is less than the mean flowrate. Between points "a" and "b" the inflow rate exceeds the controlled outflow and after "b" the inflow rate again falls below the outflow. At points "a" and "b" the two flowrates are equal. Consequently, the volume of the raw sewage requiring storage is the sum of the vertical distances between points "a" and "b" and the constant flow line. This procedure, although simple in concept, is highly dependent on the selection of the proper diurnal flow data. Diurnal flow patterns, however, vary from day to day and from season to season. The inadvertent selection of atypical diurnal patterns is likely to lead to erroneous design. Therefore, a new methodology has been developed which allows the independent assessment of the daily and the diurnal flow variations and provides for establishment of the equalization basin volume requirement on a sounder basis.

Daily flow records of eleven Ontario wastewater treatment plants were evaluated. Complete flow records for at least a two-year period were included for the statistical analysis. With the aid of a computer program, the cumulative frequency of daily mean flow was established. A typical frequency plot for a small wastewater treatment plant is shown on Figure 3a. From the plot, the maximum, the 10 percent, and the 25 percent exceedence flows can be readily established. The exceedence flow represents a particular flowrate that is exceeded at a given percent of time. A plot of the flow analysis of a large plant is shown on Figure 3b. Although the shape of the cumulative frequency curves can be dissimilar, the exceedence flow levels are obtained with ease.

The maximum, the 10 percent, and the 25 percent exceedence flowrates are plotted against the annual mean daily flowrate on Figure 4. The general equation for the best fit for the data points is the following power function:
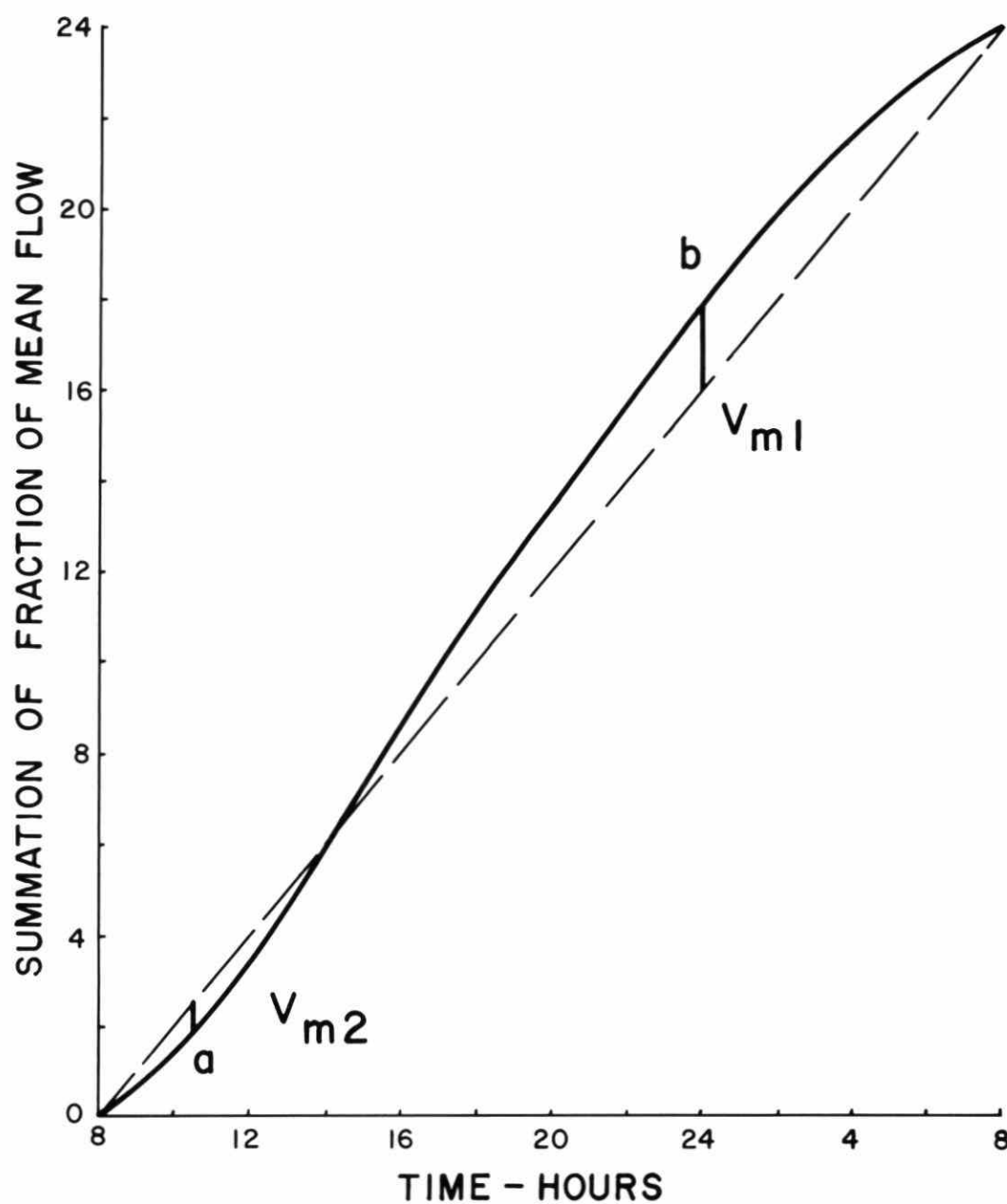
CUMULATIVE FLOW FOR 24-HOUR PERIOD

FIG. 2

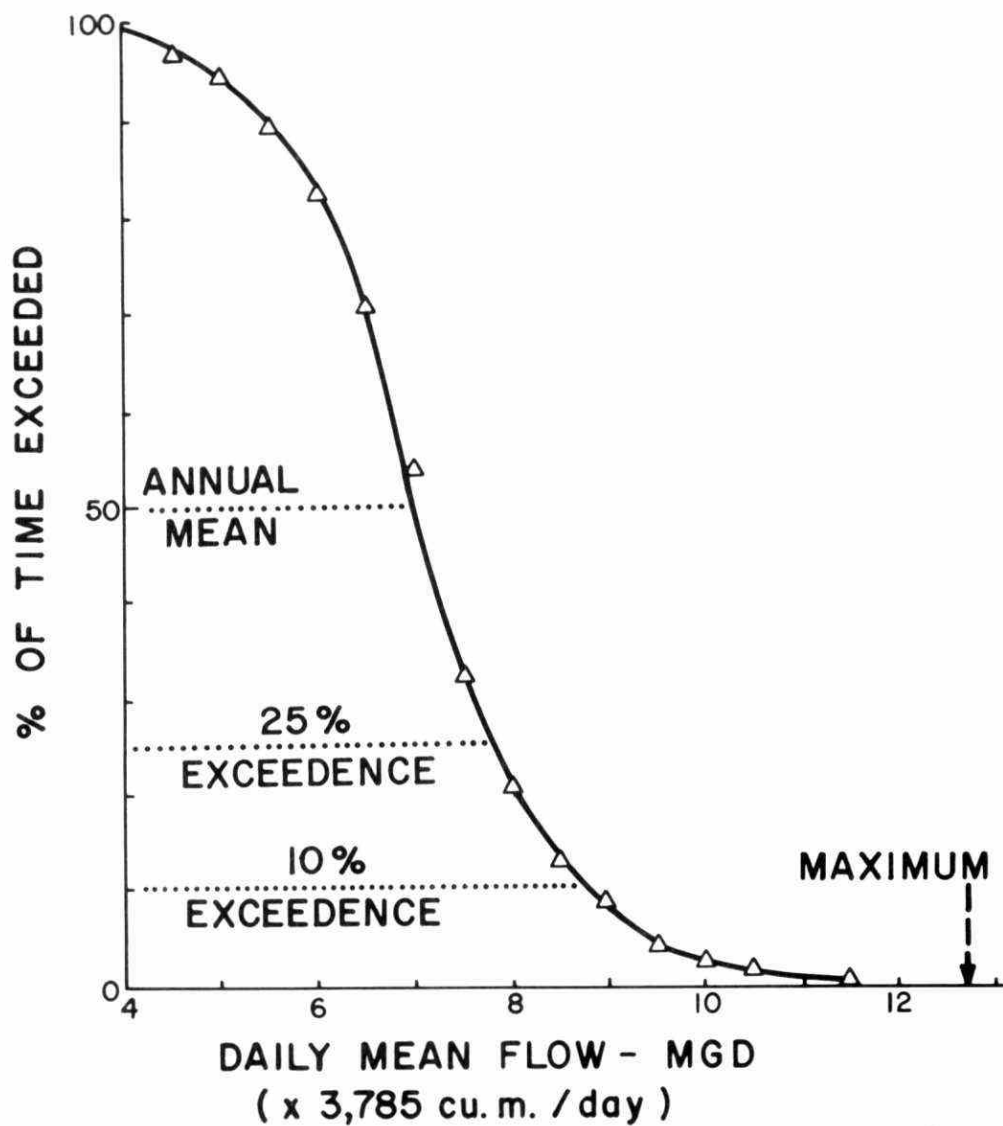FREQUENCY OF EXCEEDENCE OF DAILY MEAN FLOWS, SMALL STP

FIG. 3a

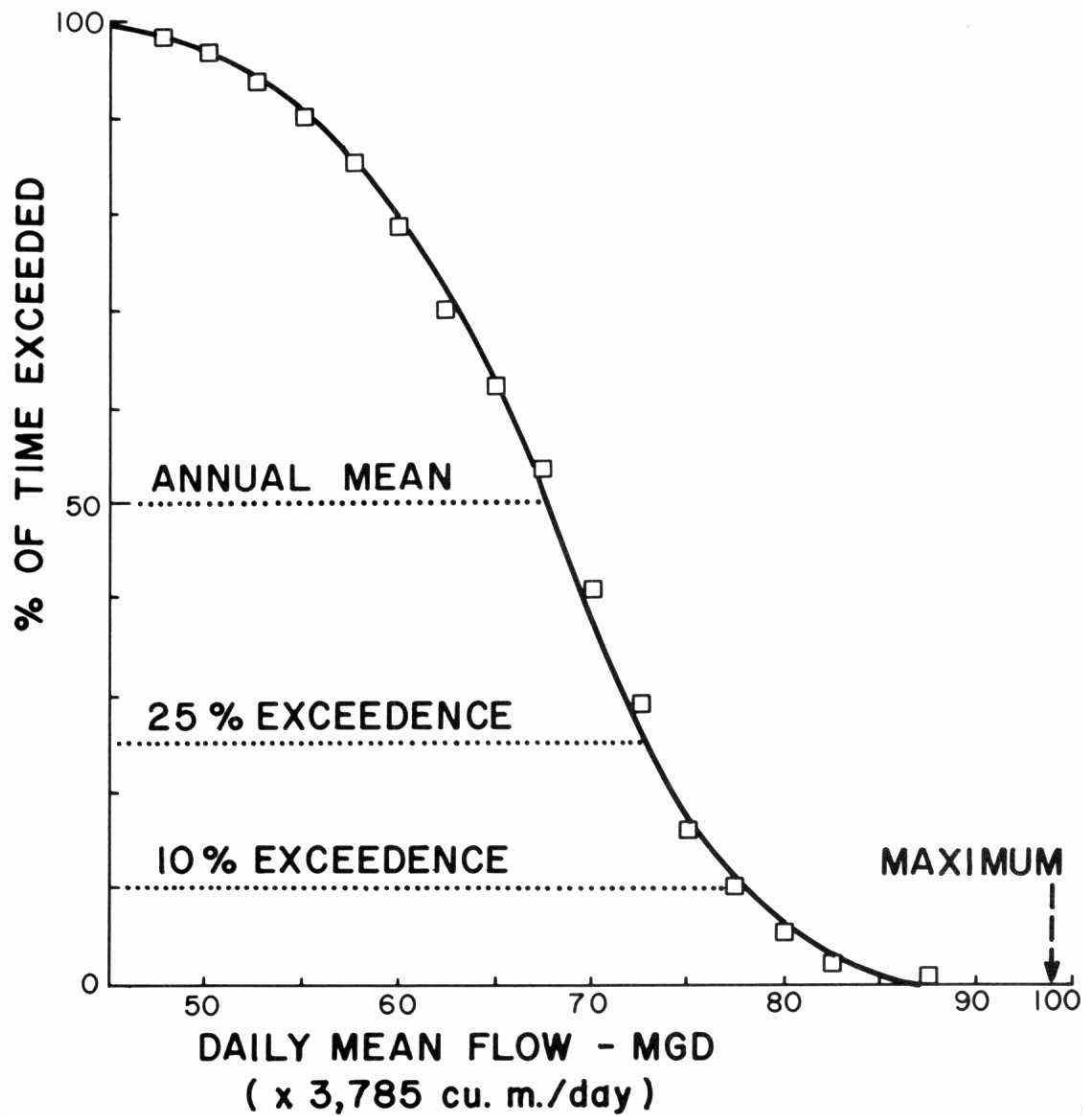FREQUENCY OF EXCEEDENCE OF DAILY MEAN FLOWS, LARGE STP

FIG 3b

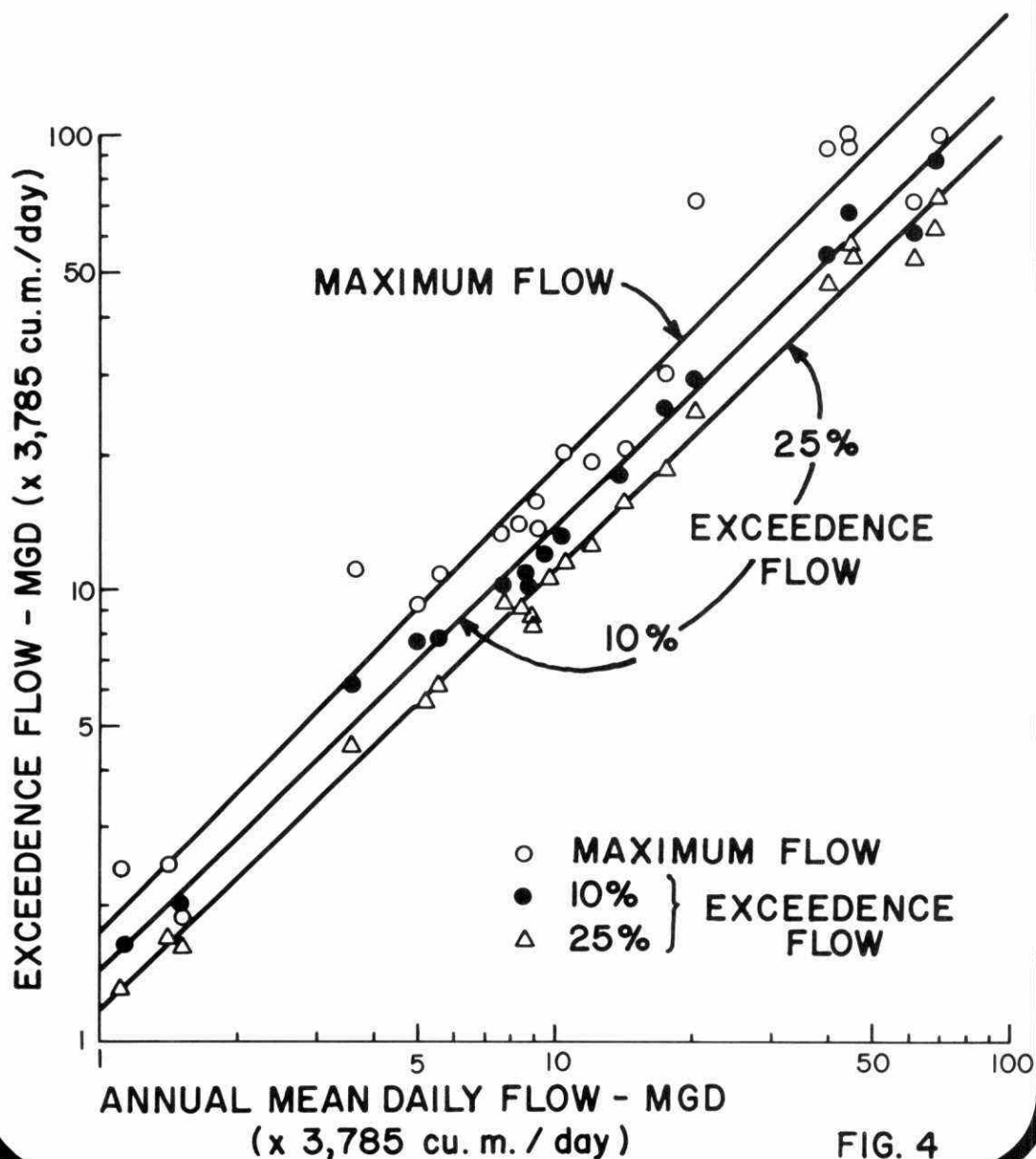THE DEPENDENCE OF THE MAXIMUM, 10 % AND 25 % EXCEEDENCE FLOWS ON THE ANNUAL MEAN DAILY FLOW

FIG. 4

$$\overline{Q}e = b\overline{Q}a^n \tag{1}$$

$\overline{Q}a$ = annual mean daily flowrate

b,n = characteristic parameters

The parameter values were obtained by regression analysis. Generally, the "n" is near unity. In this case the parameter "b" is numerically equal to the Daily Peaking Factor, F, at a particular exceedence level. The Daily Peaking Factor is, by definition, the ratio of the exceedence flow to the Annual Mean Daily Flow:

$$F = \frac{\overline{Q}e}{\overline{Q}a} \tag{2}$$

The data points, particularly the maximum flowrates, are considerably dispersed. This reflects the particular characteristics of the individual sewer systems. Three of the municipalities investigated have combined sanitary and storm sewers. It is apparent that the combined sewer system yields consistently higher Peaking Factors during wet weather. The correlation improves somewhat for the 10 percent and the 25 percent exceedence flowrates. Nevertheless, from the data analysis it can be concluded that the Daily Peaking Factors are independent of the Annual Mean Daily Flowrate.

The diurnal flow characteristics were evaluated from the daily flow charts of five wastewater treatment plants. Charts were selected for days on which the flow corresponded to the maximum day, 10 and 25 percent daily exceedence levels. These gave a wide distribution of flow conditions for each system. On Figure 5 a representative plot of instantaneous flowrate against time is shown. Diurnal flow variation results in a time period, (ΔT) when the instantaneous flowrate exceeds the daily mean. On this diagram the average flow of the peaking period together with mean Daily Flowrate are also shown. Each chart was analyzed to determine the average flowrate for several peak intervals by integrating the instantaneous flowrate. The Diurnal Peaking Factors (Fd) were obtained by normalizing the Peak Flowrates with respect to the Mean Daily Flow.

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DIURNAL FLOW VARIATION

FLOW RATE - MGD (x 3,785 cu. m./day)

AVERAGE PEAK FLOW

MEAN DAILY FLOW

STORED VOLUME

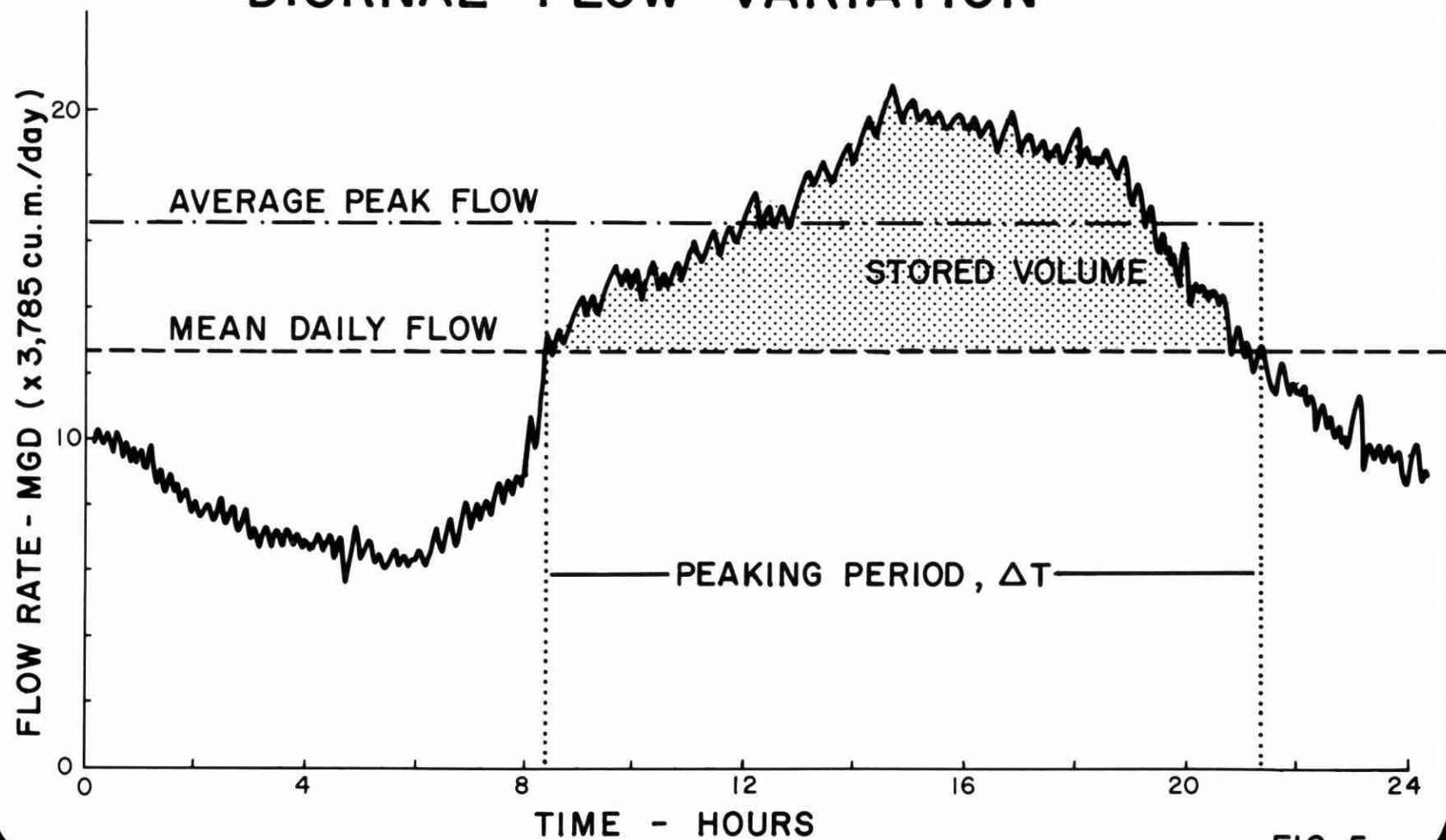PEAKING PERIOD, ΔT

TIME - HOURS

FIG. 5

The selected peak periods were 1-hour, 4-hour, 6-hour and 12-hour. The Daily Peaking Factors (F), together with the Diurnal Peaking Factors (Fd), are tabulated in Table 1. Generally, the Diurnal Peaking Factors decrease with increasing daily mean flow. Thus, maximum diurnal flow variation tends to occur when the infiltration and storm water components of the sewage are at minimum value. As expected, the Diurnal Peaking Factors decrease with increasing peaking periods. Although short-term peaking factors have no significance in the determination of the Equalization basin volume, they are essential for establishing the pumping and the flow control requirement to and from the basin. The Daily and the 12-hour Peaking Factors serve as key parameters for the realistic design of the equalization basin.

The development of the methodology is as follows. If the flow were to be equalized to maintain a mean daily flow $(\overline{Q})$, then the volume (V) of raw sewage requiring storage is the time integral of the difference between the instantaneous flowrate (Q) and the mean daily flowrate $(\overline{Q})$ over the peaking period (ΔT).

$$V = \int_{\Delta T} (Q - \overline{Q})\, dt = \int_{\Delta T} Q\, dt - (\overline{Q})\, (\Delta T) \qquad (3)$$

This volume corresponds to the shaded area in Figure 5.

The evaluation of Equation 3 requires prior knowledge of the peaking period. An examination of flow records reveals that the peaking period equals 0.5 ± 0.13 days in most cases. In terms of an average peaking period of 0.5 day (i.e. ΔT 0.5 days) and the corresponding diurnal peaking factor, Equation 3 reduces to

$$V = 0.5\, (Fd - 1)\, \overline{Q} \qquad (4)$$

The mean daily flowrate $(\overline{Q})$ is related to the annual mean flow $(\overline{Q}a)$ by the following equality:

$$\overline{Q} = F \times \overline{Q}a \qquad (5)$$

The combination of Equation 4 and Equation 5 yields an expression for the desired equalization volume as a function of the Annual Mean Daily Flow $(\overline{Q}a)$, the Daily Peaking Factor (F), and the 12-hour Diurnal Peaking Factor (Fd) as follows:

$$V = 0.5\, (F)\, (Fd - 1)\, \overline{Q}a \qquad (6)$$

AVERAGE DAILY AND DIURNAL PEAKING FACTORS

FLOW CONDITIONS	DAILY PEAKING FACTOR	DIURNAL PEAKING FACTOR			
		1hr	4hr Peaking	6hr Periods	12hr
MAXIMUM DAILY	1·7	1·22	1·19	1·18	1·13
10% EXCEEDENCE	1·3	1·20	1·15	1·14	1·14
25% EXCEEDENCE	1·12	1·24	1·23	1·20	1·16
ANNUAL MEAN	1·00	1·24	1·21	1·19	1·16

TABLE I

The estimated equalization basin volumes are given in Table 2. Full equalization is based on maximum daily flow conditions. It requires a storage volume of 11 percent of the annual mean daily flow. This represents an average value, since the parameters were derived from composite data of several plants. Some variations due to specific local sewage management are expected to exist. Nevertheless, it compares favourably with previous estimates of 12.5 percent (1) and 12 percent (7) derived by other methods.

The methodology developed also allows direct estimation of basin volumes required for partial equalization. In this case, the parameter values are based on the flow pattern at a particular exceedence level. For example, if 90 percent of the daily flow throughout the year were to be equalized, then the equalization basin would be designed on the basis of data for 10 percent exceedence flow. The interpretation of 90 percent equalization is of course, that full equalization can be achieved 90 percent of the time.

The equalization volumes given in Table 2 denote minimal volumes for typical sewage treatment plants. To account for plant to plant and annual variability, the incorporation of a safety factor is desirable. Using a safety factor of 1.25, the empirical design volumes become 14 percent, 13 percent and 11 percent of the mean annual daily flow for full, 90 percent and 75 percent equalization, respectively.

The expected utilization of the equalization basins's capacity under various mean daily flow conditions is illustrated on Figure 6. The plot corresponds to full equalization. The basin is filled to full capacity only under maximum flow conditions. During annual mean flow conditions 72 percent of the capacity is utilized. A Daily Peaking Factor of 0.5 corresponds to minimum daily flow. In this case, approximately 36 percent of the basins's capacity is utilized for eliminating the diurnal fluctuation.

COST BENEFITS

A comparison of the cost of treatment facilities with and without equalization was developed. Equalization basin costs depend on the size, as well as the methods and materials of construction. New

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THE DEPENDENCE OF THE EQUALIZATION BASIN VOLUME ON % EQUALIZATION

$$V = (\Delta T) \times F \times (Fd - 1) \times \bar{Q}_a$$

% EQUALIZATION	ΔT	F	Fd	NOMINAL EQUALIZATION VOLUME (% of annual mean daily flow)
100	0·5	1·7	1·13	11· 05
90	0·5	1·3	1·14	9· 10
75	0·5	1·12	1·16	8·96

TABLE 2

THE DEPENDENCE OF
STORED SEWAGE VOLUME
ON THE DAILY FLOWRATE
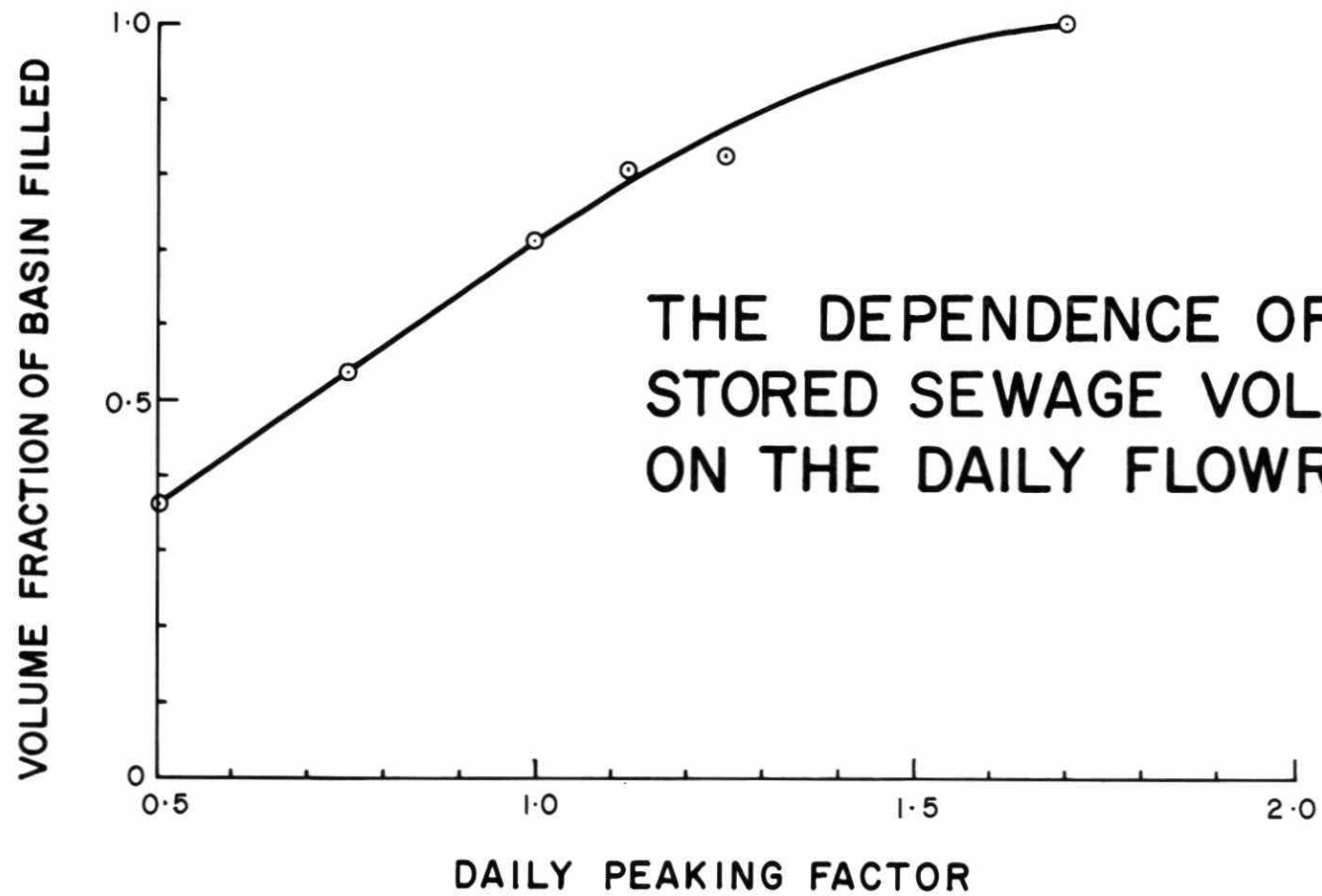
FIG. 6

basins may be built of earth, concrete, or steel. Normally earthen basins are the least expensive, but may be subject to erosion if proper precautions are not taken in the design. Concrete basins require considerably less land area; therefore, they are preferred when the available space is limited.

Successful performance requires adequate aeration and mixing. Previous studies have recommended both agitation and aeration of proposed equalization basins to prevent solids settling and odour formation (6, 8, 9, and 10). For this paper diffused aeration has been assumed to provide both mixing and odour control and was based on normal aeration requirements of 1.25 to 2.0 cfm (cubic feet per minute) per 1,000 gallons of equalization tank volume. Properly designed equalization basin facilities also include provisions for sewage pumping, automatic flow measurement, and flow regulation.

The estimated cost of a concrete basin, including diffused aeration, is presented on Figure 7. The costs were derived from the U.S. Environmental Protection Agency's cost estimate for wastewater treatment facilities, dated January 1971. The cost figures were updated using the ENR Construction Cost Index for Toronto, Ontario, to July 1974.

Equalization of the diurnal fluctuations in sewage flows is believed to have two effects. It improves the performance of all processes that are sensitive to changes in hydraulic loadings and it decreases the size of all treatment stages designed on the basis of peak loadings. To illustrate the relative costs associated with plants designed to treat equalized and variable flow with no flow bypassing in each case, the following alternatives are considered:

1. Variable Flow
2. Fully Equalized Flow
3. Partly Equalized Flow

The relative cost of wastewater treatment facilities with and without equalization for plants of 5 and 50 MGD mean annual daily flow are shown in Table 3. The cost evaluation for the equalized wastewater treatment plants was based on predicted peak hydraulic loads.
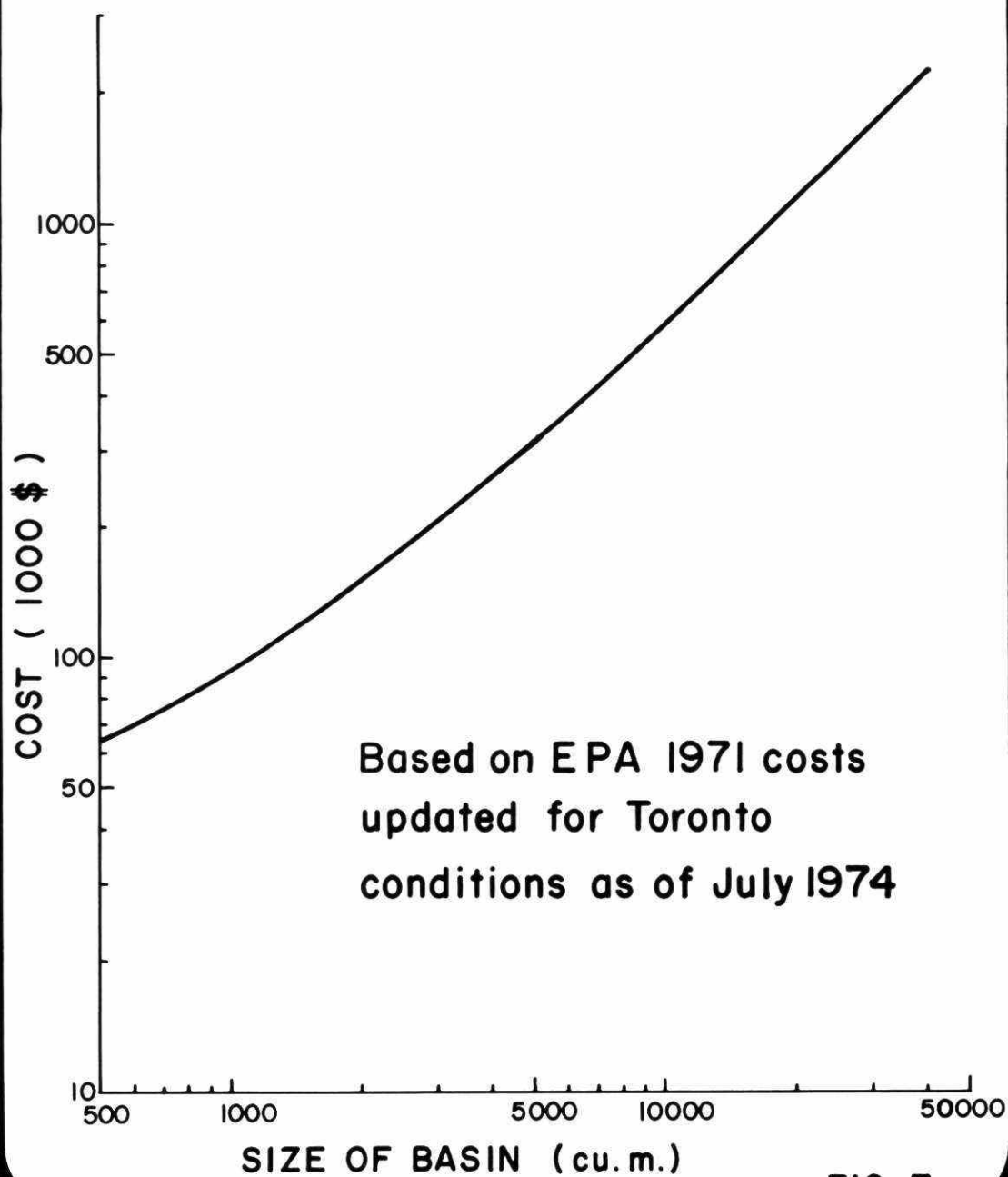
COST OF
EQUALIZATION FACILITIES

Based on EPA 1971 costs
updated for Toronto
conditions as of July 1974

SIZE OF BASIN (cu. m.)

FIG. 7

RELATIVE COST OF
WASTE WATER TREATMENT FACILITIES

	5 MGD PLANT	50 MGD PLANT
NO EQUALIZATION	*100·0*	*100·0*
75% EQUALIZATION	**98·5**	**98·7**
90% EQUALIZATION	*96·4*	*96·4*
100% EQUALIZATION	*95·5*	*94·7*

TABLE 3

From the tabulated data, it is apparent that the introduction of equalization results in small reductions in construction costs. A 5 percent cost benefit may be achieved by full equalization in the 5 MGD and the 50 MGD plants. The major cost savings are due to a reduction in the size of primary and secondary sedimentation tanks. Partial equalization does not appear to be attractive economically. The saving in size reduction is negated by the incremental cost of the equalization basin.

CONCLUSIONS

To establish meaningful criteria for designing equalization facilities, extensive analysis of daily and diurnal variations of flow in Ontario sewage treatment plants was carried out. The flow variations were expressed in terms of daily and diurnal peaking factors.

Significant conclusions of the flow variation analysis were:

- The peaking factors are independent of plant size.
- Variation in peaking factors reflect differences in plant operation. The most important contribution to variability is the combination of raw sewage with storm runoff.

Utilizing the daily and the diurnal peaking factors, a methodology was developed for sizing of equalization basins. The volume of the basin was estimated to be 11 percent to 14 percent of the annual mean flow. The volume size insures continuous operation without set-backs due to overflow.

A cost comparison for two plant sizes (5 MGD and 50 MGD) indicated that design for flow equalization conditons may result in a small construction cost reduction over variable flow design.

The assessment of flow equalization practices yielded inconclusive evidence concerning effluent water quality improvement. On the basis of limited data, process improvements due to equalization may be realized in the performance of primary and secondary wastewater treatment plants. However, further research is warranted to evaluate the effect of flow equalization on the performance of these units.

REFERENCES

1. Colson, N.V., "Characterization of Urban Land Runoff", Paper presented at the American Soc. Civ. Engr. Nat. Water Resources Conference, Los Angeles, Calif. (1974).

2. Storage and Treatment of Combined Sewer Overflows, EPA Series R2-72-070 (1972).

3. Click, C.N., and Mixon, F.O., "Flow Smoothing in Sanitary Sewers", J. Water Poll. Control Fed., 46, 522 (1974).

4. Nelson, O.F. and R. Agnew, "Quick Treatment of Combined Sewer Overflow Shows Promise", Water and Wastes Engineering, 10, 39 (1973).

5. Flow Equalization, EPA Technology Transfer Seminar Publication, Washington, D.C. (1973).

6. Process Design Manual for Upgrading Existing Wastewater Treatment Plants, U.S. EPA Technology Transfer, Oct. 1974.

7. Smith, R., R.G. Eilers and E.D. Hall, "Design and Simulation of Equalization Basins", National Environmental Research and Development Centre Report, EPA-670/7-73-046 (1973).

8. James F. MacLaren Limited, "To Establish Viable Methods of Maintaining Waste Treatment Facility Efficiencies with Reference to Waste Variations", Contract No. OGR2-0363, Ministry of the Enviroment, Etobicoke, Ontario.

9. LaGrega, M.D. and Keenan, J.D., "Effects of Equalizing Wastewater Flows", J. Water Poll. Control Fed. 46, 123 (1974).

10. LaGrega, M.D., "A Study of the Effects of Equalization of Wastewater Flows", Ph.D. Dissertation, Syracuse University, 1972.

11. Thomann, R.V., "Variability of Waste Treatment Plant Performance", J. of Sanitary Eng. Division, ASCE, 813 (1970).

12. Foess, G.W., Meenahan, J.G., and Blough, D.B., "Evaluation of In-line and Side-line Flow Equalization Systems, Paper presented at 48th WPCF Conference, Miami Beach, Fla. (1975).

UPGRADING PRIMARY CLARIFIER PERFORMANCE
BY CHEMICAL ADDITION

by

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CONCEPTS

It will be useful to recall a few points important to the clarification of degritted sewage in primary settling tanks.

- the concentration of suspended solids is sufficiently low, so that unhindered settling of particles occurs;

- solids suspended in sewage do not settle by simple gravity separation; flocculation plays an important part, whether chemicals are added or not;

- the settling characteristics of the suspension and the hydraulic efficiency of the clarifier must be considered in a performance evaluation of the tank;

- the main parameters for measuring the performance of a clarifier are suspended solids and BOD (or organic suspended solids) removal; and,

- overflow rate and detention time are the parameters used in the design of primary clarifiers.

The performance of primary clarifiers can be improved in a number of ways, which can be used singly or in combination. They include:

- provison of equalization tanks to achieve more uniform hydraulic loading;

- special arrangements (inlet and outlet devices, baffles, etc.) to increase the hydraulic efficiency of the tank;

- use of more efficient settling devices, such as tube settlers, tray separators, etc.; and,

- use of chemicals to increase flocculation and thereby separation.

This paper deals with the last mentioned method. The recent requirement for phosphorus removal from sewage has resulted in the widespread use of chemical precipitation in sewage treatment. Increased solids and BOD reductions have been reported concurrent with phosphorus reduction (Weston, 1971). This paper reports the effects of chemical addition on clarification at two primary treatment plants (Sarnia and Windsor) practising phosphorus removal. Also presented are some of the results of a pilot study at the Canada Centre for Inland Waters, Burlington. The research was carried out over a period of two and a half years. Complete results have been previously reported by Heinke (1973) and Heinke, Qazi and Tay (1975).

CURRENT DESIGN GUIDELINES

The guidelines of a few agencies on the design of primary clarifiers, with and without the use of chemicals, are presented.

ASCE and WPCF "Sewage Treatment Plant Design" (1959):

Page 93:   "Primary Tanks - Surface settling rates not followed by secondary treatment shall not exceed 600 USgpd per sq. ft. for plants having a design flow of 1.0 mgd or less. Higher rates may be permitted for larger plants."

page 92:   "Practical liquid depth of 7 to 12 ft. combined with area provide the nominal detentions of 1 to 2 hrs. for primary tanks."

page 111:   "When chemicals are used, the primary tanks generally have a minimum detention period of 2 hrs. and a settling rate of no more than 1000 USgal per sq. ft. of area per day based on average flow."

Ontario Ministry of Environment (unpublished, currently under review):

	PRIMARY TREATMENT PLANT	SECONDARY TREATMENT PLANT		
	PRIMARY SETTLING TANK	PRIMARY SETTL. T.		SECONDARY SETTL. T.
		NO WASTE RETURN SLUDGE	WASTE RETURN SLUDGE	
HYDRAULIC LOADING OVERFLOW RATE, GPD/FT^2 AT AVG. FLOW*		650 - 1000	500 - 650	NO CHEM. 800 CHEM. ADD. 650-700
AT PEAK FLOW**	NO CHEM. 800-1000 LIME UP TO 1000 ALUM, FERRIC 650-800 CHL.			
DETENTION TIME, HRS. (THEOR.) AT PEAK FLOW	MIN. 1 HR. (DEPTH 10 - 15 FT.)	MIN. 1 HR. (DEPTH 10 - 15 FT.)		
SOLIDS LOADING: lb/ft^2/DAY				UP TO 50 AT PEAK FLOW

*   AVG. FLOW ≡ AVERAGE DAILY DRY-WEATHER FLOW

** PEAK FLOW ≡ MAX. HOURLY FLOW, NORMALLY TAKEN AS 3X AVG. FLOW

European Practice (Kalbskopf, 1970):

Detention Time	0.5 to 2.5 hrs.
Overflow rate:	
horizontal flow tanks	0.5-1.5 m/h (240-720 Igal/ft^2/day)
vertical flow tanks	2.5 m/h (1200 Igal/ft^2/day)
Mean flow velocity	0.8-1.5 cm/sec.

No information is available for the case of chemical addition, but recognition of the favourable effect of flocculation on settling is noted.

A primary clarifier treating domestic sewage without chemical addition is generally able to achieve 30 to 60% suspended solids removal. Weston (1971) concluded from several plant studies in which polymers were used that the average value of suspended solids removal increased from 38% with no polymer to 65% with polymer use. Lime addition at

three plants produced solids removals from 75 to 90%. Voshel and Sak (1968) reported the successful use of organic polymers to increase suspended solids removal, as shown in Figure 1. Several other studies have obtained similar results.

There has been no attempt in this study to obtain a wide cross-section of current design guidelines on the effect of chemical addition. The limited sample indicates that present guidelines generally do not recognize increased settleability of chemical flocs by allowing higher overflow rates or lower detention times. However, guidelines are under review and may be changed in the near future, when studies, such as this one, provide information on which to make informed decisions.

LABORATORY AND PLANT STUDIES

The performance of a settling tank depends on two things: the settling characteristics of the suspension, and the hydraulic efficiency of the tank. Settling column tests can be used to study the settling characteristics of a suspension in the laboratory. Tracer studies and velocity (and suspended solids concentration) measurements in the tank are useful to assess the hydraulic efficiency of the tank. The performance of the settling tank may be predicted by the combination of information obtained in the laboratory and in the plant.

A number of plants were investigated for possible inclusion in the study. The prime requirement was that chemical addition at the primary stage had to be practised over the length of the study (1972-1975). Only the Sarnia and Windsor S.T.P. met this condition. Pilot plant work at CCIW, Burlington was added later.

Laboratory studies

139 settling column tests (in addition to 50 tests on Toronto sewage), were preformed at Sarnia, Windsor and Burlington under the following chemical conditions:

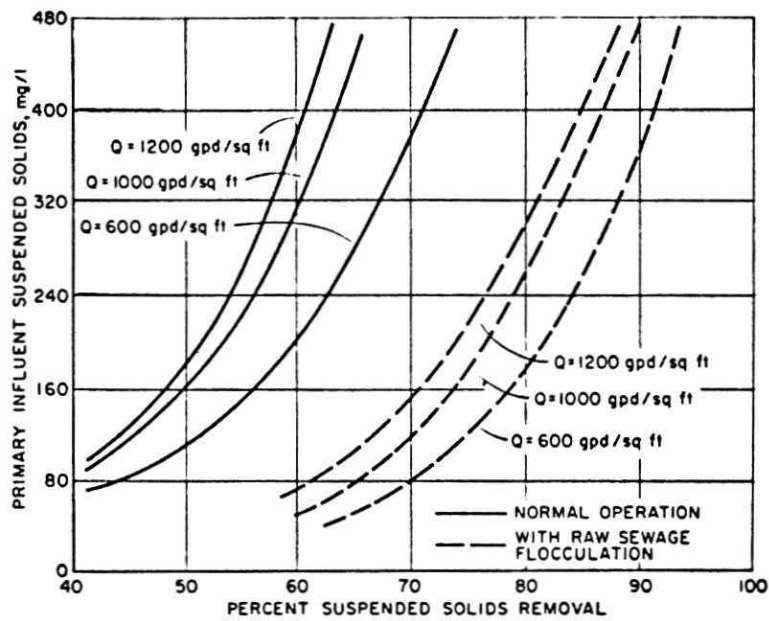|	
No chemical addition	(50 tests)
Ferric chloride addition	(20 tests)
Alum addition	(12 tests)
Ferric chloride plus polymer (Dow Purifloc A23) addition	(41 tests)
Alum plus polymer addition	(16 tests)

Figure 1. INFLUENCE OF OVERFLOW RATE AND PRIMARY SS ON PERCENT
SS REMOVAL (Voshel and Sak, 1968)
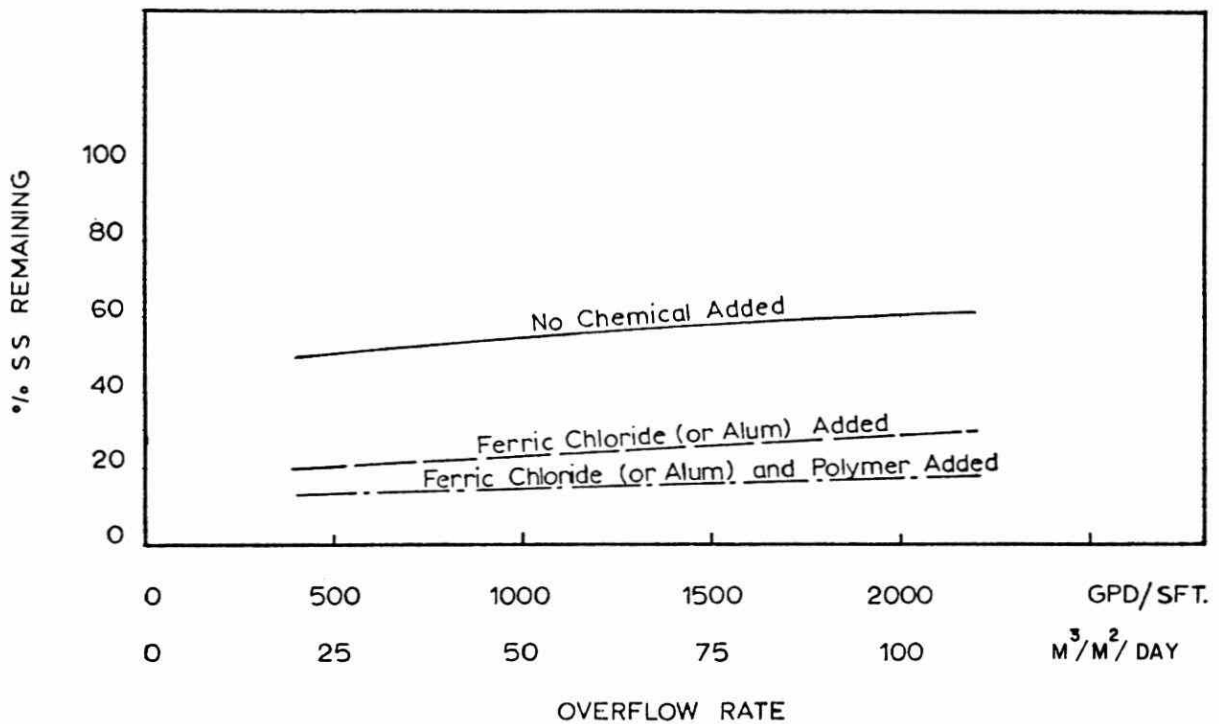(USgpd/sq ft x 0.408 = $m^3/m^2/day$)

Figure 2. EFFECT OF OVERFLOW RATE ON CLARIFICATION EFFICIENCY -
TYPICAL RESULTS FOR DOMESTIC SEWAGE

Fifty-nine tests were performed at Sarnia, 70 at Windsor and 10 at Burlington. Analysis showed that the settling behaviour of flocs at all three plants for the same chemicals was quite similar. Furthermore the performance of alum and ferric chloride flocs was about the same. Consequently, the results of all tests have been summarized under the following three categories:

- Settling characteristics of flocs - no chemical addition

- Settling characteristics of flocs - alum or ferric chloride addition

- Settling characteristics of flocs - alum or ferric chloride and polymer additions

The settling characteristics of physical-chemical flocs were examined in relation to the following parameters. It is realized that there are interdependencies:

(a) Overflow rate

(b) Mixing

(c) Detention time

(d) Settling depth

(a) Effect of Overflow rate

Figure 2 shows the effect of overflow rate on clarification efficiency for domestic sewage (as % SS remaining), with and without chemical addition, after 60 minutes of settling under quiescent conditions. It shows that, for a given detention (settling) time (60 minutes in the case shown), overflow rate has a small effect on clarification of flocculent suspensions. From settling column tests, it is estimated that if the overflow rate increases from 500 to 2000 gpd/sq ft (or 25 to 100 m^3/m^2/day) the suspended solids in the effluent will increase by about 10% with no chemical addition, 9% with ferric chloride or alum addition, and 5% with ferric chloride or alum and polymer additions. Percentages stated refer to influent suspended solids concentration.

(b) Effect of Mixing

In Sarnia and Windsor, chemicals were added at a location where reasonable mixing conditions existed (in a pre-aeration tank at Sarnia, and a grit removal tank or at a raw sewage pump at Windsor) but which provided only a few minutes of mixing/flocculation time, particularly after the polymer was added and before the suspension was discharged into the settling tank. The samples for the settling column tests were taken just before the inlet to the tank and poured into the settling column initially without any further mixing.

The settling performance predicted by the settling column compared very closely with that of the settling tank at low overflow rates (about 500 gpd/sq ft or 25 m^3/m^2/day), when a scale-up factor of 2 was applied to detention time as commonly practised. At higher overflow rates, however, the settling tank produced much better effluent quality than predicted by the settling column. This lack of comparison at higher overflow rates finally led us to believe that the suspensions taken for settling tests were not fully flocculated, that further mixing/flocculating was occurring in the tank and that the flocs would grow when subjected to further mixing before the settling column test. Therefore, a series of tests was conducted in which suspensions with chemical addition were slowly mixed for 20 minutes by a propeller-type mixer before settling in the column. This series of tests showed significant improvement in settling rate due to additional mixing (see Figure 3 curves A and B). These tests were carried out in Windsor only, but probably the general findings are applicable to Sarnia as well.

Figure 3 (curves A and B) shows that the settling rate of mechanically flocculated suspensions was much faster that that of the unstirred suspensions. It should be noted that this difference diminishes with increase in detention time. The extent of improvement in settleability will depend heavily on the degree of flocculating potential existing in the suspension. If the flocs are already fully developed, slow mixing may not have any beneficial effect and curves A and B in Figure 3 will tend to coincide. On the other hand, if the suspension
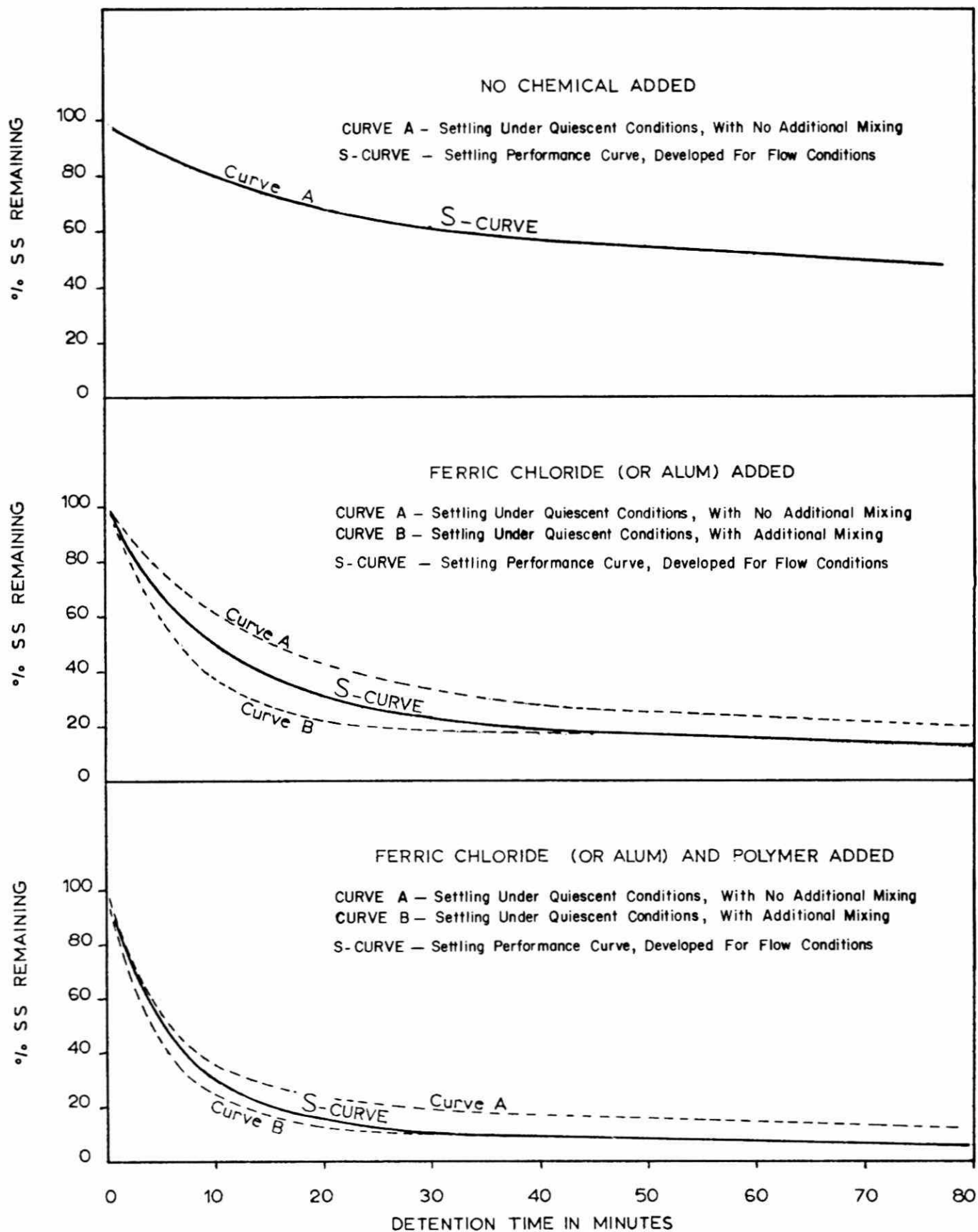
Figure 3. EFFECT OF MIXING AND PERFORMANCE CURVES

was mixed prior to the settling tank for a short time only, then slow mixing will be beneficial to floc formation and subsequent faster settling. Further work in this area is planned.

(c) Effect of Detention Time

Figure 3 also shows the effect of detention time, with partial flocculation (curve A) and with complete flocculation (curve B). The curves are fairly steep in the initial range of detention time, and become progressively flatter with passage of time. When flocs were fully developed, most of the settling occurred in about 45 minutes without chemical addition and in about 20 minutes with chemical addition (curve B). When flocs were not fully developed (curve A) clarification required a little longer, (45 minutes for ferric chloride or alum only, and about 30 minutes in combination with polymer).

If detention time is increased beyond what is described above, only a small improvement in effluent quality will be achieved, as evident from the flat curve. On the other hand, if the detention time is reduced much below the stated times, drastic deterioration of effluent quality will result. It is, therefore, very important that, for effective performance of settling tanks, all of the wastewater flow should be receiving the appropriate minimum detention times indicated. Discussion on the performance curves (S-curves) shown is delayed until later.

(d) Effect of Settling Depth

Figure 4 shows the effect of settling time on clarification efficiency, with and without chemical addition, with settling depth as a parameter. The curves represent the average over the depth shown. For a given detention time, clarification deteriorates only slightly as settling depth increases. The effect of depth further diminishes with an increase in detention time, particularly with chemical addition. With ferric chloride or alum and polymer additions, depth of settling ceases to have any significant effect on settling efficiency after about 30 minutes of detention time. This may mean that only small improvement in clarification efficiency can be expected from the use of trays or tubes, when chemicals are used.
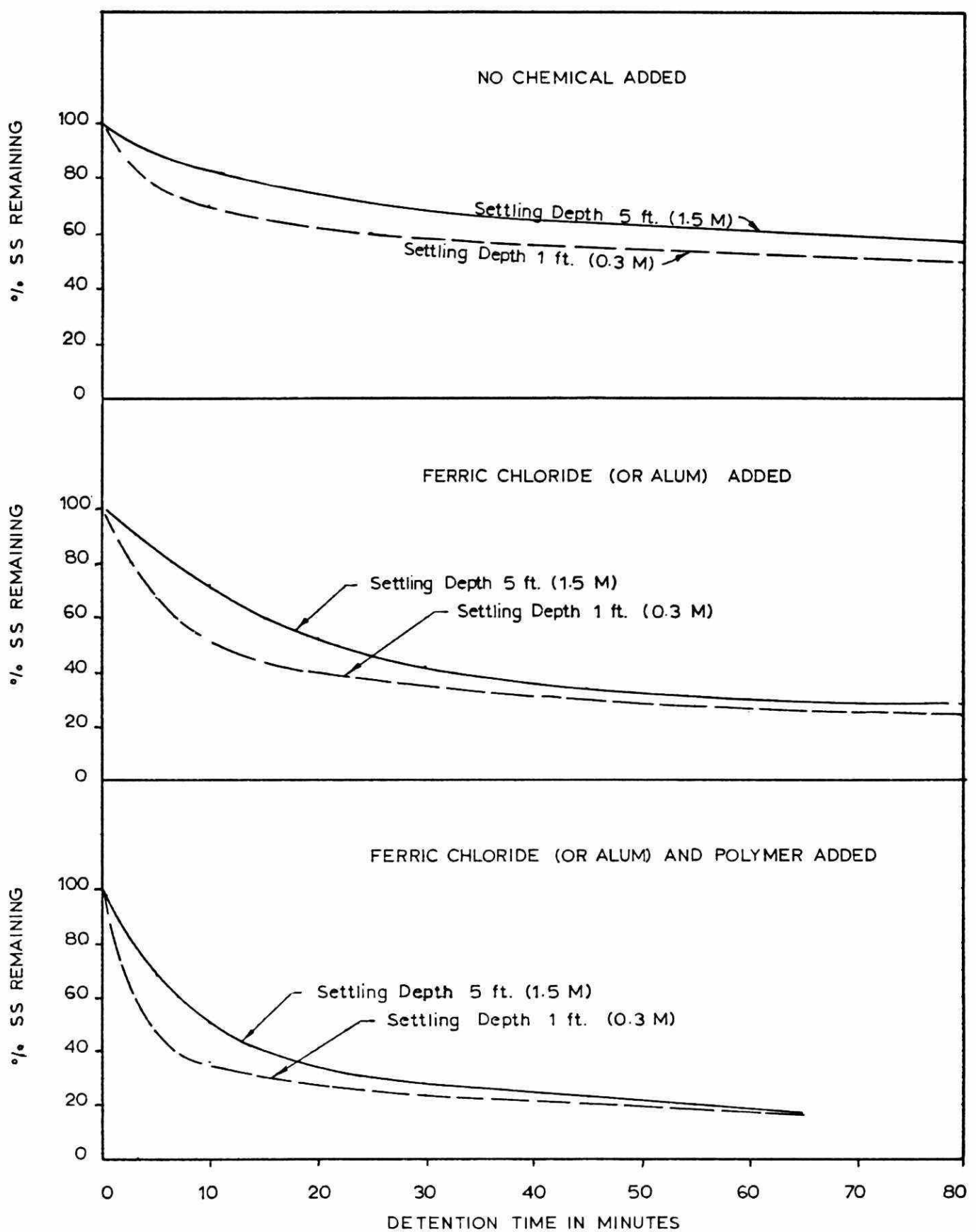
Figure 4.   EFFECT OF SETTLING DEPTH ON CLARIFICATION EFFICIENCY

In summary, laboratory studies of settling tests on domestic sewage, with and without chemical addition, have shown that the most important parameter for efficient settling is effective detention time. The influence of either overflow rate or settling depth alone is minor. Chemical addition greatly increases clarification. Comparison of results on effluent quality predicted from laboratory settling tests, applying accepted scale-up factors, and actual plant performance is good. Mixing of suspensions prior to settling is beneficial to faster settling.
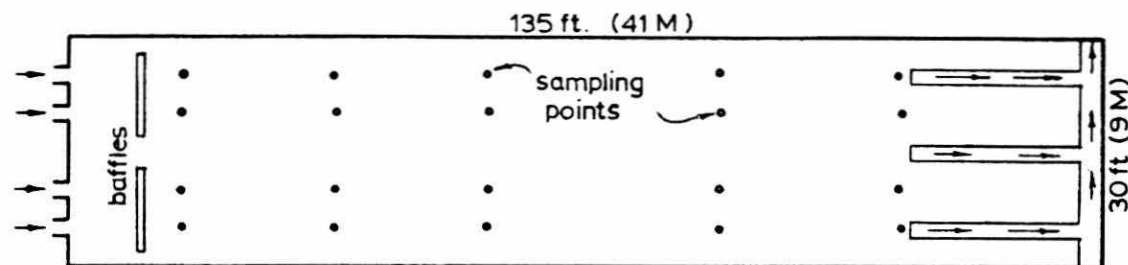
Plant Studies

Information on physical features, flow rate and influent and effluent quality (with and without chemical treatment) for the Sarnia, Windsor and Burlington plants are given in Figures 5, 6, and 7. The Sarnia (rectangular tanks) and Windsor (circular tanks) clarifiers are typical of horizontal flow clarifiers, whereas the Burlington pilot plant is typical of a vertical flow clarifier. The detailed analysis and discussion of the performance of the three plants has been presented by Heinke (1973) and Heinke, Qazi and Tay (1975). Summary results on suspended solids for Sarnia and Windsor are shown in Figures 8 and 9. The conclusion can be drawn that:
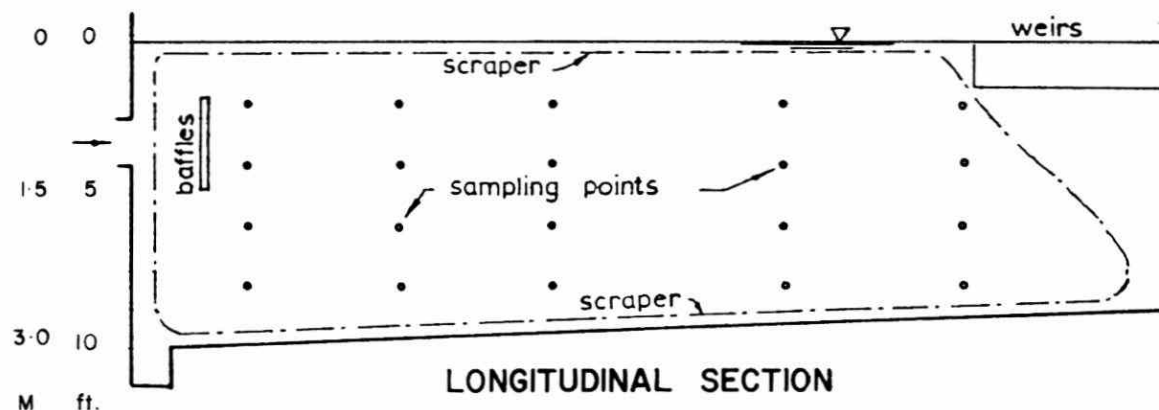
- The Sarnia S.T.P. was an efficient primary treatment plant. Effluent quality was further substantially improved by the addition of chemical treatment.

- The Windsor S.T.P. did not produce a good primary effluent. However, substantial improvement in effluent quality was achieved through chemical addition.

- The Burlington pilot plant received a much stronger sewage than the Sarnia or Windsor plant. On a percentage removal basis the plant's performance was somewhere between that of Sarnia and Windsor. Chemical treatment produced a great improvement in effluent quality.

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PHYSICAL FEATURES

No. of Settling Tanks	4	
Length	135 ft	40.5m
Width	30 ft	9.0m
Depth	9 ft	2.7m
Volume	36,500 cu ft	1035m³
Weir Length	162 ft	49.5m
Weir Loading	12,300 gpd/ft	185m³/m/day
Outlet Trough	3	
Length	22.5 ft	6.8m
Width	2 ft	0.6m
Det. Time	2-3/4 hr	

135 ft. (41 M)

30 ft (9 M)

baffles

sampling points

PLAN

weirs

scraper

baffles

sampling points

scraper

LONGITUDINAL SECTION

INFLUENT (1971 and 1972)

Parameters	Average	Range	
		Max.	Min.
Flow Rate, MGD	6.35	12.00	5.25
m³/day	2.88 x 10⁴	5.45 x 10⁴	2.38 x 10⁴
Overflow Rate, gpd/ft²	390	2090	326
m³/m²/day	19	102	16
Suspended Solids, mg/l	102	608	8
BOD, mg/l	88	182	52
Total P, mg/l	5.7	22.6	2.2

EFFLUENT (1971 and 1972)

Chemical Addition	SS mg/l			BOD mg/l			Total P mg/l		
	Avg.	Max.	Min.	Avg.	Max.	Min.	Avg.	Max.	Min.
None	38	102	4	63	152	33	5.2	8.0	1.0
Ferric Chloride	23	74	9	34	80	17	0.9	5.3	0.2
Ferric Chloride plus A-23	12	54	4	31	59	25	0.8	2.9	0.1
Alum	37	49	20	40	61	27	1.5	3.3	0.4
Alum plus A-23	30	73	19	37	55	27	1.4	2.9	0.2

Chemical Dosages: Ferric chloride 10-20 mg/l, Alum 80-100 mg/l, Polymer 0.3-0.5 mg/l

Figure 5. SARNIA TREATMENT PLANT

PHYSICAL FEATURES

No. of Settling Tanks	4	
Surface Diameter	120 ft	36m
Surface Area	11,300 ft^2	1020m^2
Side Wall Depth	11 ft	3.3m
Volume	121,800 ft^3	3450m^3
Weir Length	357 ft	107.1m
Weir Loading	16,800 gpd/ft	250m^3/m/day
Det. Time	3.1 hr	
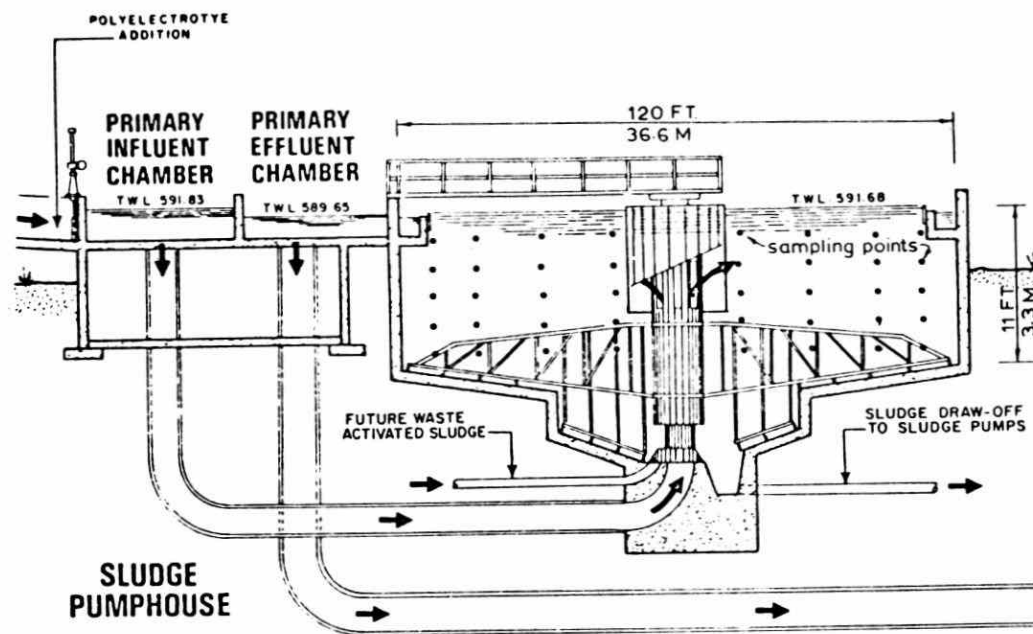
PRIMARY
SETTLING TANKS

INFLUENT (1971 and 1972)

Parameters	Average	Range	
		Max.	Min.
Flow Rate, MGD	16.0	72.5	5.2
m^3/day	7.25 x 10^4	33.0 x 10^4	2.4 x 10^4
Overflow Rate, gpd/ft.^{2}	350	1600	115
m^3/m^2/day	18	80	6
Suspended Solids, mg/l	102	1054	26
BOD, mg/l	101	346	27
Total P mg/l	5.1	9.5	2.0

EFFLUENT (1971 and 1972)

Chemical Addition	SS mg/l			BOD mg/l			Total P mg/l		
	Avg.	Max.	Min.	Avg.	Max.	Min.	Avg.	Max.	Min.
None	61	309	22	76	297	21	4.8	9.3	1.9
Alum Only (5-23 May/72)	42	129	8	36	90	12	1.9	5.9	0.3
Alum plus Polymer (24 May - 16 July/72)	30	60	6	36	77	14	1.1	2.6	0.3
Ferric Chloride plus polymer (22 Dec/74 - 11 Jan/75)	36	73	21	50	120	21	0.7	1.6	0.4

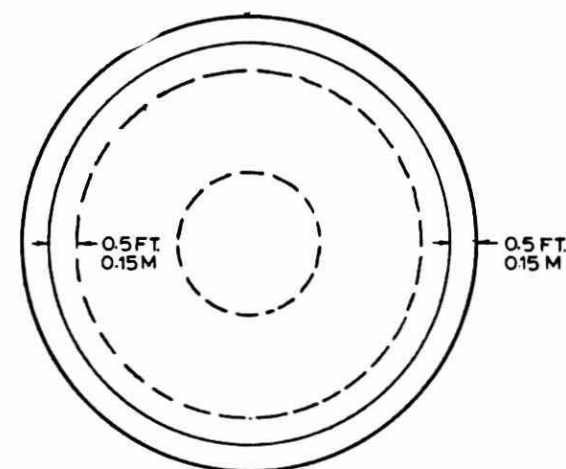Chemical Dosages: Alum 80-108 mg/l, Ferric Chloride 17 mg/l, polymer 0.3-0.4 mg/l

Figure 6.   WINDSOR TREATMENT PLANT

PHYSICAL FEATURES

No. of Settling Tanks	1	
Diameter	8 ft	2.6m
Surface Area	50 ft²	5.4m²
Ave. Depth	7 ft	2.1m
Volume	2192 gal	10.0m³
Weir Length	22 ft	6.7m
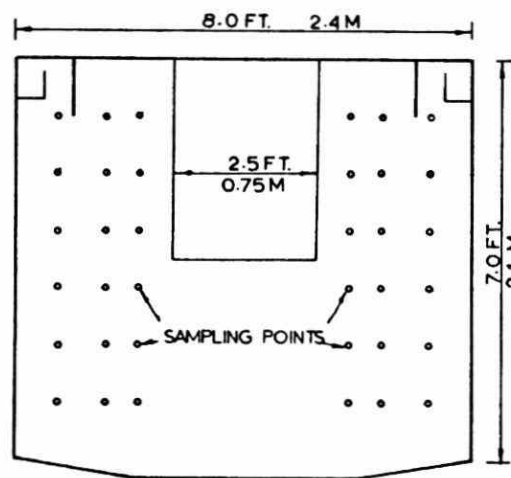
ELEVATION

PLAN

INFLUENT

Parameters	Average	Range	
		Max.	Min.
Overflow Rate gpd/ft.²	600	600	600
m³/m²/day	30	30	30
Suspended Solids, mg/l	220	480	118
BOD, mg/l	134	274	46
Total P, mg/l	6.2	8.7	4.0

EFFLUENT

Chemical Addition	SS mg/l			BOD mg/l			Total P		
	Avg.	Max.	Min.	Avg.	Max.	Min.	Avg.	Max.	Min.
None	104	134	84	93	114	78	4.4	4.9	3.3
Ferric Chloride	55	72	26	49	56	41	1.2	1.7	0.4
Ferric Chloride plus Polymer	31	52	14	32	38	24	1.2	1.7	1.0

Chemical Dosages: Ferric Choride 18 mg/l, Polymer 0.3 mg/l

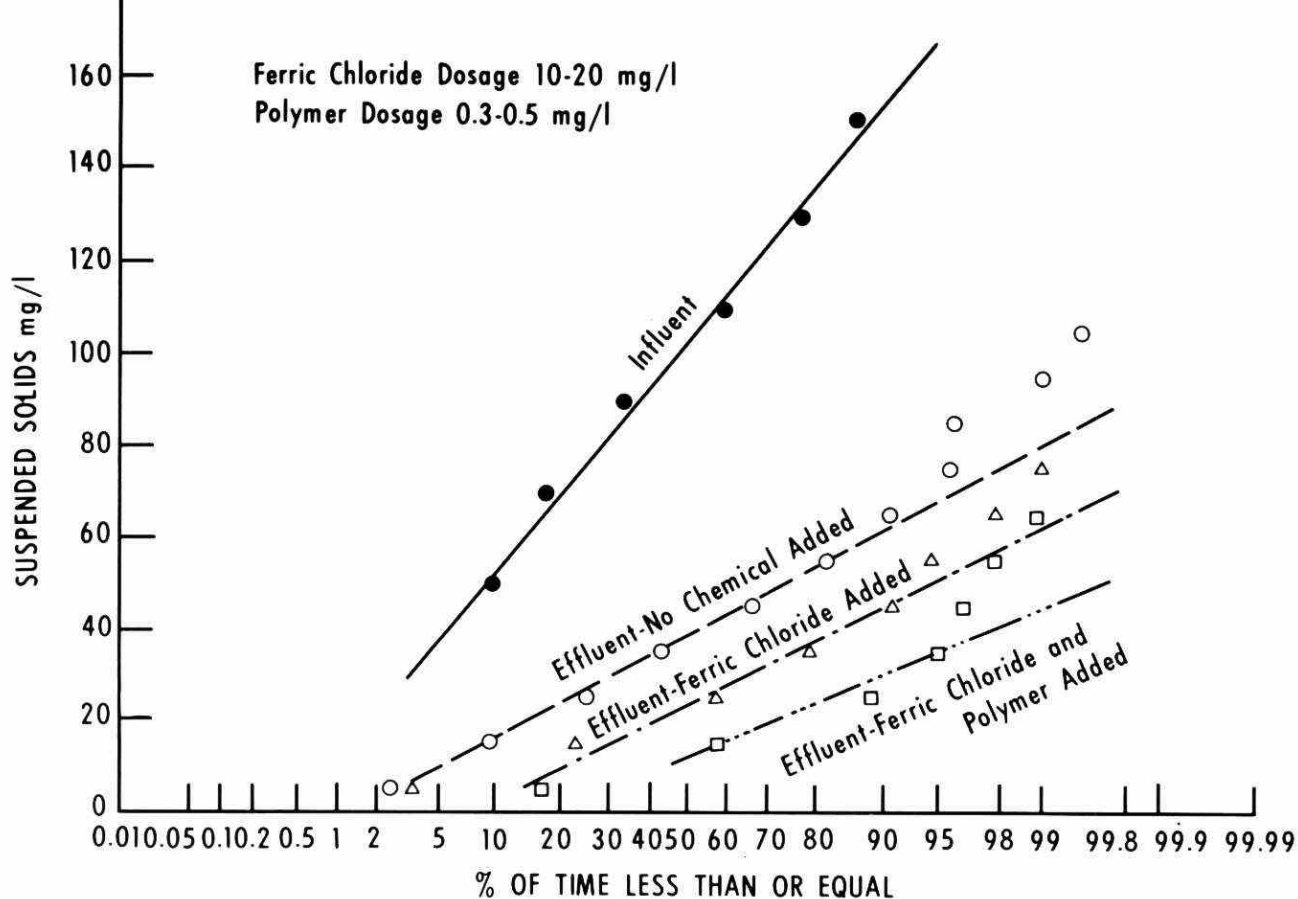Figure 7. CIWW PILOT PLANT, BURLINGTON

Figure 8   PERFORMANCE OF TREATMENT PLANT IN SARNIA
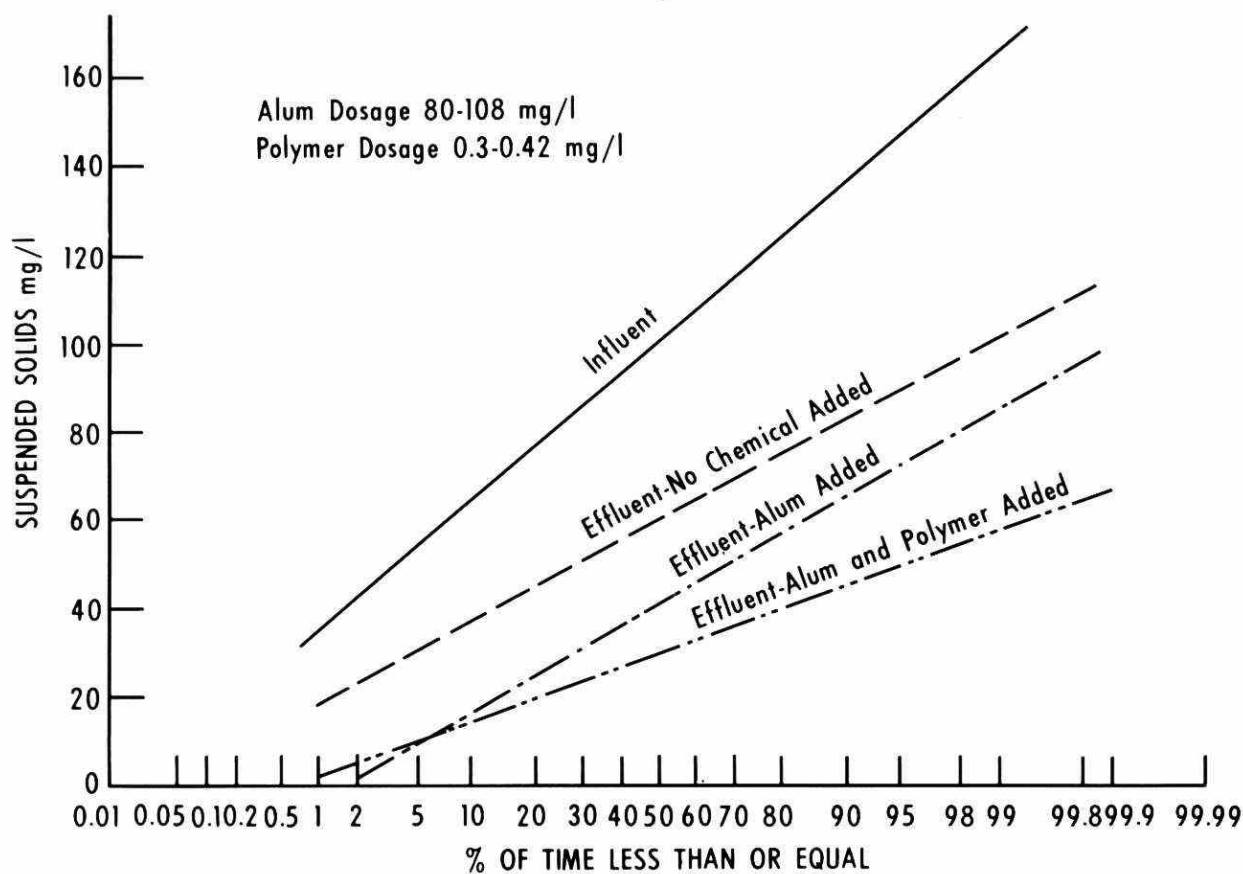

Figure 9   PERFORMANCE OF TREATMENT PLANT IN WINDSOR

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Tracer Studies. Tracer studies were used to determine actual detention times of the settling tanks under study. The pulse technique was used. A known quantity of Rhodamine B was dumped at the clarifier inlet and dye concentrations were recorded continuously at the effluent weir by a Turner fluorometer. Initial experiments were carried out to establish that the results obtained in later work were representative of the overall plant.

About 100 dye tests were carried out. Summary results are presented in Table 1. Typical C-curves are shown in Figure 10. Definition of symbols is as follows:

t_i -   time interval for initial indication of the tracer in the effluent, in minutes

t_p -   time to reach peak or maximum concentration, in minutes

t_g -   time to reach centroid of the curve (actual mean detention time), in minutes

T -   theoretical detention time, in minutes

e -   hydraulic efficiency, t_g/t

c, c_o -   dye concentration at time t and t_o

Discussion on Actual Detention Time. As defined above, actual mean detention time of flow is determined by the value of t_g (centroid of C-curve). For an ideal settling tank (which by definition has no "dead" zones), the value of t_g/T will be unity. For an actual tank, it will be less than unity because of the presence of "dead" zones and, therefore, reduced effective volume of settling tank.

The results in Table 1 show that the rectangular tank at Sarnia is much closer to ideal (average $t_g/T = 0.73$) than the circular tank in Windsor (average $t_g/T = 0.36$). This means that in the rectangular tank, about 3/4 of the total volume is effective, whereas in the circular tank only 1/3 of the volume is effective. The results further show that the t_g/T ratio increases, although to a small extent, with increase in overflow rate. For the pilot circular tank at Burlington, however, the value of t_g/T varied greatly (0.38 to 0.73) as the overflow rate increased from 600 to 2000 gpd/sq ft (from 30 to 100 $m^3/m^2/day$).

TABLE 1

Hydraulic Efficiency Parameters

A. AT SARNIA

Overflow Rate		Parameters in Minutes				$c = t_g/T$	t_i/T
gpd/sq.ft.	m^3/m^2/day	t_i	t_p	t_g	T		
500	25	38*	60*	105*	160*	0.66	0.24
1000	50	24	39	62	80	0.78	0.30
1500	75	19	31	41	53	0.77	0.36
2000	100	16	25	29	40	0.73	0.40

B. AT WINDSOR

Overflow Rate		Parameters in Minutes				$c = t_g/T$	t_i/T
gpd/sq.ft.	m^3/m^2/day	t_i	t_p	t_g	T		
500	25	20	46	60	200	0.30	0.10
1000	50	10	25	33	100	0.33	0.10
1500	75	7	18	27	67	0.40	0.10
2000	100	5*	15*	21*	50*	0.42	0.10

C. AT BURLINGTON

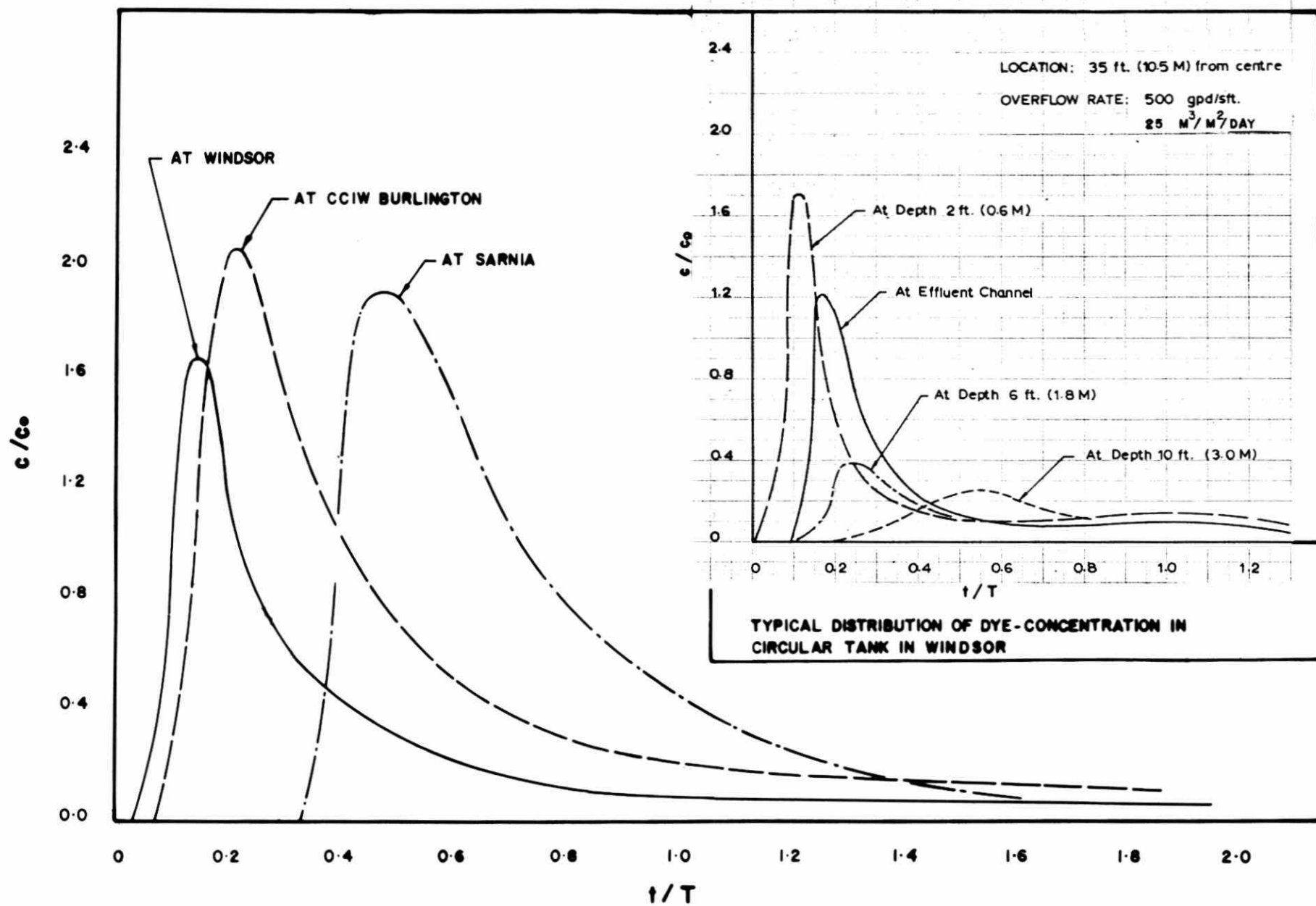| Overflow Rate		Parameters in Minutes				$c = t_g/T$	t_i/T
gpd/sq.ft.	m^3/m^2/day	t_i	t_p	t_g	T		
600	30	8	28	42	110	0.38	0.07
1000	50	5	25	36	66	0.55	0.08
1500	75	--	--	--	---	----	----
2000	100	5	16	24	33	0.73	0.15

* Obtained by extrapolation

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FIGURE 10: TYPICAL C - CURVES

Similarly, the comparison of other parameters between Sarnia and Windsor plants shows that the dye front (t_i) and peak (t_p) appear much earlier at Windsor than Sarnia (see Figure 10 and Table 1). The value of t_i/T, for example, is higher (0.24 - 0.40) for the Sarnia tank than Windsor (0.1), whereas the Burlington tank falls in between (0.07 - 0.15), in the range of overflow rates of 500 - 2000 gpd/sq ft (25 - 100 m^3/m^2/day) used in the study.

Discussion of Flow Pattern in Circular Tank. The analysis of C-curves indicates the presence of acute short-circuiting and dead zones in the circular tank in Windsor. To investigate this further, the pattern of flow was studied by tracing the dye concentrations at various depths and locations in the tank. The relative concentration and amount of dye picked up (over a sufficiently long period of time) from various depths at a given location was assumed to be indicative of the relative intensity and amount of flow passing through various depths at that location. The results of these tests showed that in the circular tanks in Windsor the flow generally occurred through the top half of the tank leaving the bottom half ineffective (see Figure 10, top right). In some cases the flow pattern was reversed and flow went to the bottom and flowed along the bottom of the tank leaving the top half ineffective.

In Sarnia no such tests were performed but it can be assumed that the flow in the rectangular tank in Sarnia was stable and uniform with few dead regions, as indicated by various hydraulic parameters.

In summary, it can be stated that the rectangular tanks in Sarnia are closer to an ideal settling basin and much superior in performance to the circular tanks in Windsor. The circular tanks in Windsor suffer drastically from the presence of dead zones.

Velocity and Suspended Solids Profiles. These tests were conducted to gain a better understanding of the sedimentation process in actual tanks under continuous flow conditions. The main objectives were to study:

- distribution of velocity, flow paths and dead zones;
- distribution of suspended solids;

- effect of overflow rate on velocity and suspended solids distribution; and,
- effect of chemical addition on settling rates in settling tanks.

Velocity and suspended solids concentration measurements were taken at various locations and depths as shown in Figures 5, 6 and 7. In Sarnia and Windsor velocity was measured by a <u>non-directional</u> Low Velocity Current Meter, built by Ontario Hydro. In Burlington, a <u>directional</u> meter (Electromagnetic Water Current Meter, 600 Series Velmeter, Model P, manufactured by Cushing Engineering Inc. U.S.A.) and two-axes automatic recorder (Watanabe Instrument Corporation, Tokyo, Japan) were used for velocity measurements. Suspended solids measurements were taken by Suspended Solids Monitor manufactured by Par-Tech.

Typical results of the study are summarized in Figures 11, 12 and 13. It should be noted that the elevations of the tanks shown are not in their true proportions and therefore the profiles for Sarnia and Windsor look steeper (and for Burlington flatter) than they really are.

<u>Discussion on Velocity Profiles</u>. It is difficult to illustrate flow patterns from nondirectional velocities (Sarnia and Windsor). The magnitudes of the velocities shown are the result of velocity components in three principal directions. The major velocity component was in the plane shown. With this caution the velocity profiles shown in Figures 11, 12 and 13 indicate that:

(i) In Sarnia, velocities were highest at about the mid-depth of the tank and most of the flow passed through the sections between 4 to 7 ft (1.2 to 2.1 m) from the surface of the water.

(ii) It is difficult to determine the flow pattern from the velocity profiles measured at Windsor. In part this may be caused by unsymmetrical results, caused by weir elevation differences. Some of the ineffective areas of the tank show surprisingly high velocities. Tracer studies indicate a "closed loop" water movement in these "stagnant" areas.
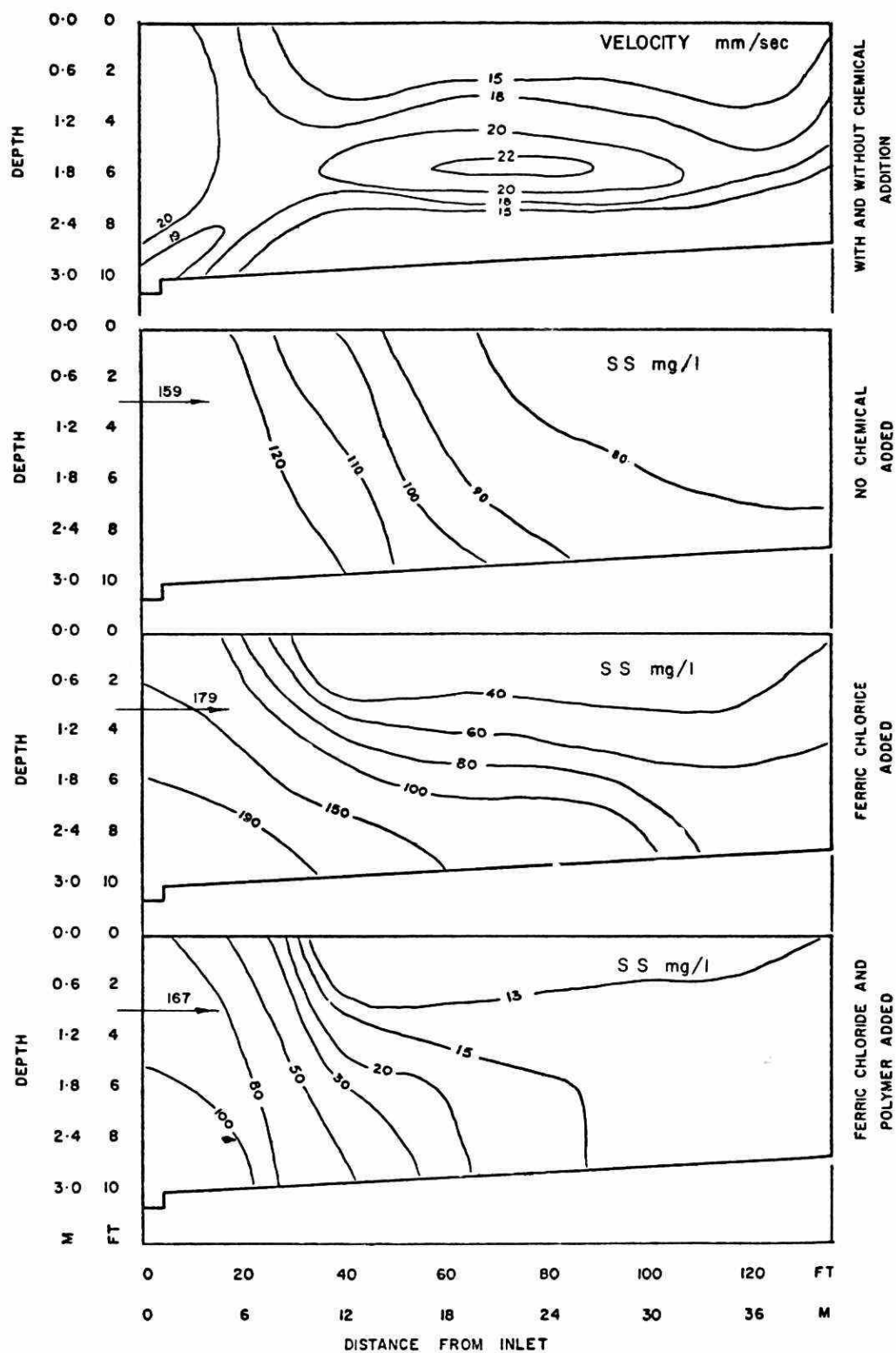
Figure 11.   VELOCITY AND SUSPENDED SOLIDS PROFILES AT SARNIA
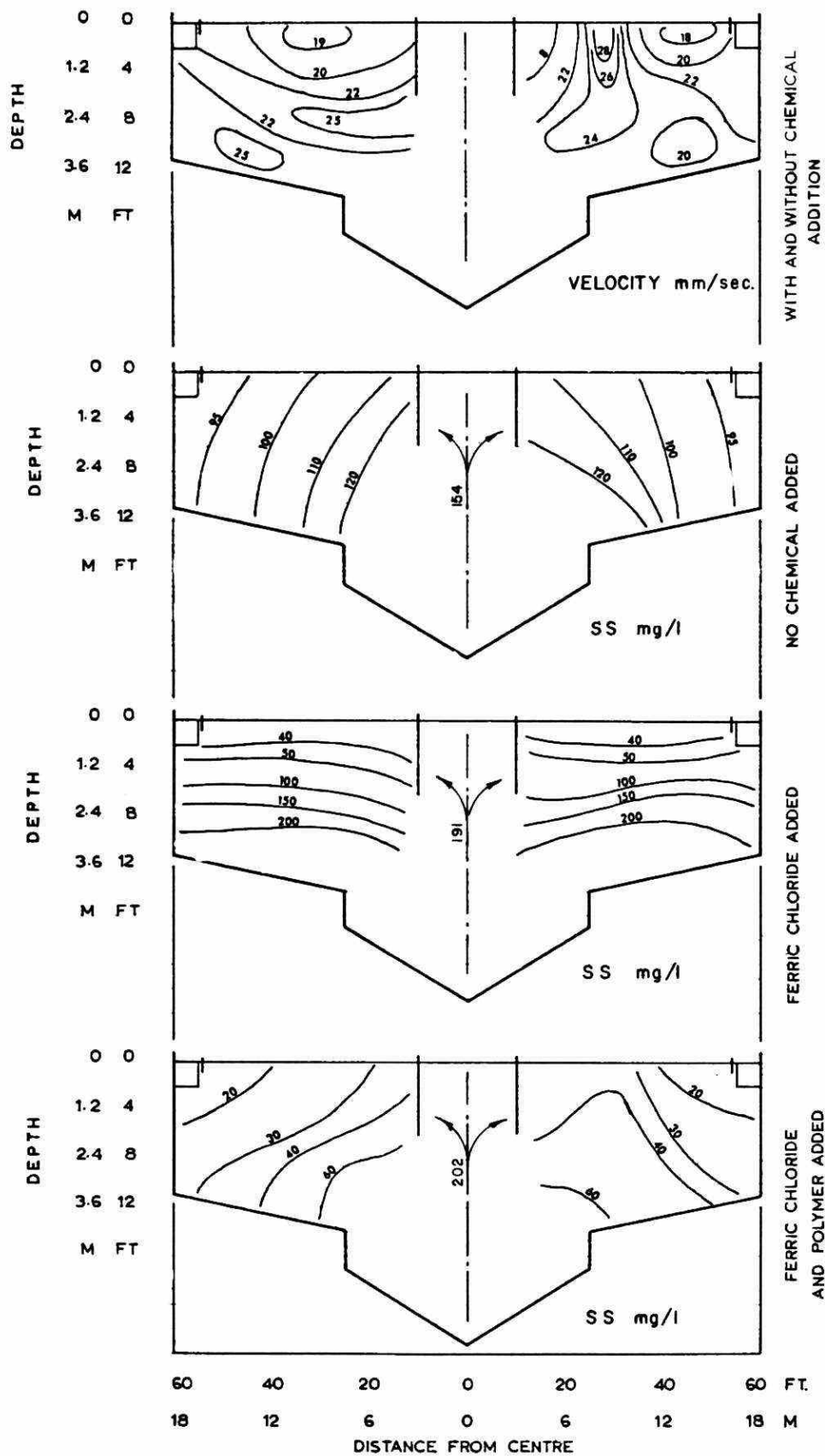Overflow Rate: 1200 gpd/sq ft (60 m³/m²/day)

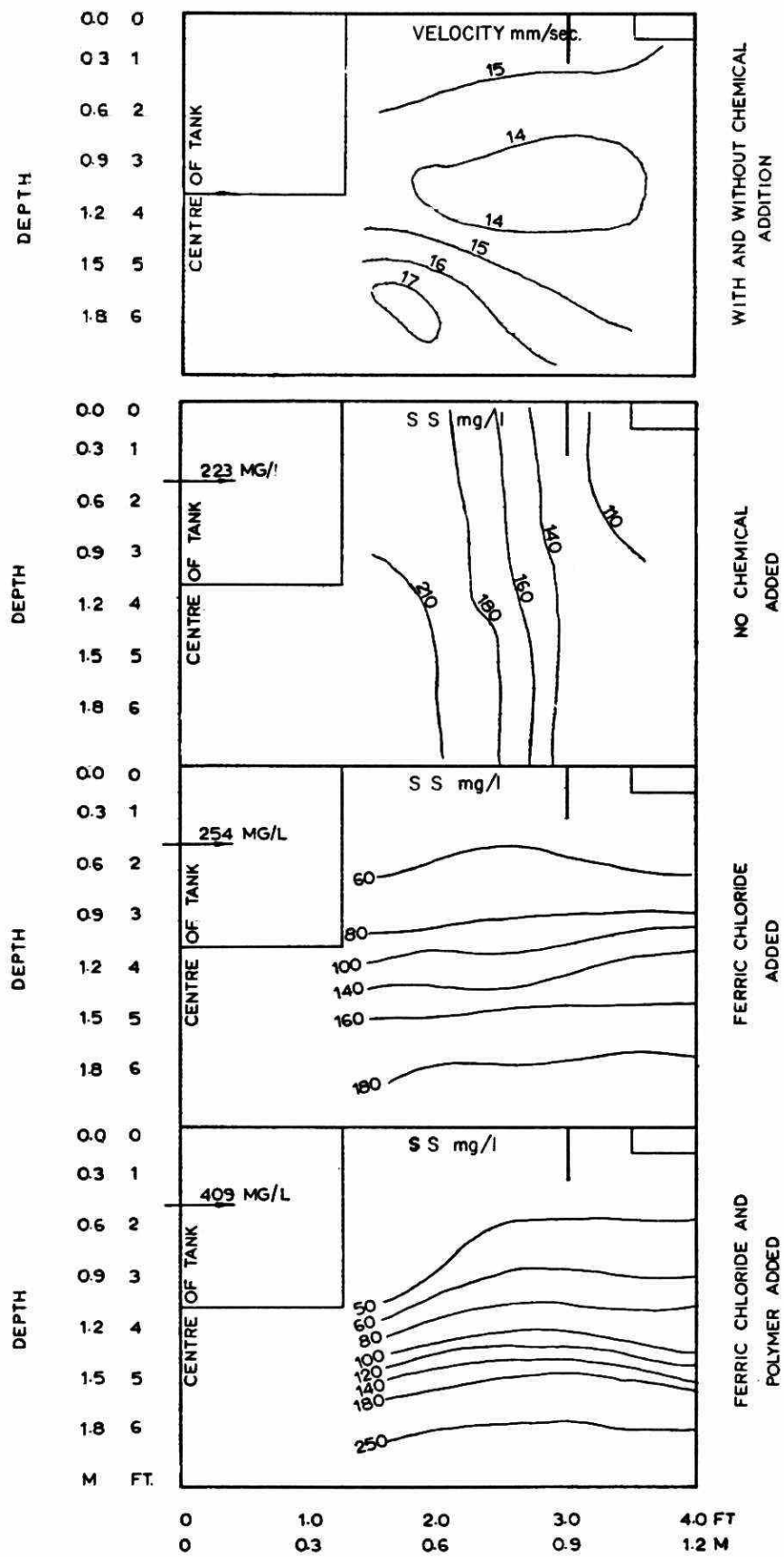Figure 12.   VELOCITY AND SUSPENDED SOLIDS PROFILES AT WINDSOR
Overflow Rate: 1000 gpd/sq ft (50 m³/m²/day)

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Figure 13. VELOCITY AND SUSPENDED SOLIDS PROFILES AT BURLINGTON
Overflow Rate: 1000 gpd/sq ft (50 m³/m²/day)

(iii) For comparison purposes resultant velocities are shown for Burlington. All three velocity components (horizontal, vertical and tangential) were in general of about the same magnitude.

Discussion on Suspended Solids Profiles. The suspended soild profiles in Figures 11, 12 and 13 show that:

(i) The addition of chemical shifted the profiles towards the inlet. In other words, when chemicals were added, the settling rate increased and clarification was accomplished within a shorter length of the tank and therefore within a shorter detention time. Best results were achieved when both ferric chloride and polymer were added.

(ii) The addition of chemicals affected the shape and slope of profiles, particularly in the circular tanks in Windsor and Burlington. With chemical addition, the profile tended to become horizontal.

(iii) The results also show (not shown in figures) that as flow rate increased (up to 2000 gpd/sq ft or 100 m^3/m^2/day as overflow rate), suspended solids profiles shifted towards the outlet as follows:

- significantly when no chemical was added;
- to a small extent when ferric chloride was added; and,
- insignificantly when ferric chloride and polymer were added.

DEVELOPMENT OF SETTLING MODEL

A number of mathematical and/or empirical models have been developed which are based on the "Theory of settling of discrete particles" and/or laboratory scale experimental results. A few researchers have introduced parameters to evaluate the hydraulic efficiency of settling tanks and their effect (in relative terms) on sedimentation. Fielder and Fitch (1959) used dye-tests results as a measure of "detention efficiency" and developed a mathematical relation-

ship to express sedimentation of flocculating suspensions. Villemonte et al (1967) developed a criterion called "critical settling depth" by matching the settling rate of suspensions at various depths and the removal performance of laboratory model of settling tank using aluminum hydroxide flocs. Clements (1966) and Clements and Khattab (1968) introduced a term "time ratio" (the ratio of effective flow-through time to effective settling time) to assess the detention efficiency of settling tanks. The flow-through time is determined from the velocity cross-sections and profiles of flow in the tank, and it is different from the one measured by dye-test. Prive (1974) supported Clement's work. Voshel (1968) developed an empirical relationship among suspended solids removal, feed loading and overflow. In our work , an attempt is made to combine the information on settling characteristics of sewage (obtained from settling tests) with the information on hydraulic characteristics of a settling tank (obtained from tracer studies or from velocity measurements) in an empirical equation, which will predict the effluent suspended solids concentration.

For efficient clarification, flocs should be fully developed through mixing ahead of the settling zone. The settling test should be carried out on a suspension simulating plant conditions. This may or may not require additional mixing ahead of the settling column test. Settling performance curves (S-curves) drawn on Figure 3 are believed to represent plant conditions. They can be represented by a mathematical expression:

$$SS_{removed} = S_o - S = k \, St^n \qquad \text{Equ. 1}$$

$$\text{or} \qquad S = \frac{S_o \, A}{t^n + A} \qquad \text{Equ. 2}$$

where

S_o = Suspended solids in influent to settling tank, mg/l

S = Suspended solids remaining in effluent of settling tank, mg/l

t = Detention time in minutes

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k = Flocculation constant

A = 1/k

n = Settling constant

The values of constants for the suspensions under study (domestic wastewater) were determined by "least-square fit".

Chemical Addition	N	K	A = 1/k
No Chemical	0.64	0.067	15.0
Ferric Chloride (or Alum)	0.90	0.133	7.5
Ferric Choride (or Alum) plus Polymer A23	0.93	0.286	3.5

The work on predicting the hydraulic performance of a settling tank from velocity and suspended solids concentration profiles is incomplete at this time. Further work is now under way. It is possible to predict the effluent suspended solids concentration by superimposing the C-curve (from tracer studies) onto the S-curve (from settling tests) and carrying out a fractional integration. The area under the C-curve defines the flow and the area under the S-curve gives suspended solids concentration. If most of the C-curve falls over the flat portion of the S-curve, then the concentration of suspended solids remaining in effluent can be read off the S-curve at the centre of gravity of the C-curve, i.e., at t_g (actual mean detention time).

The effluent quality, as predicted from the settling model, was compared with the plant effluent at Sarnia, Windsor and Burlington. Table 2 shows a typical comparison between the performance of the actual tanks and the model, using actual mean detention time (50 minutes) as a parameter. There is close agreement. It should be noted that, although the overflow rates and theoretical detention time were different at the three plants, the effluent quality of these plants remained about the same as long as the actual mean detention time was the same. The general validity of the model for the settling of domestic sewage must be verified at other plants. The procedure should be applicable for any suspension.

TABLE 2

COMPARISON OF SETTLING MODEL AND REAL TANK PERFORMANCE

Treatment Plant at	Detention Time		Overflow Rate		Influent to Settling Tank mg/l	Effluent, SS Remaining	
	Act. Mean	Theor.	gpd/ft^2	m^3/m^2/day		Real Tank	Model
With No Chemical Addition							
Sarnia	50	68	1200	60	159	53%	55%
Windsor	50	165	600	30	154	55%	55%
Burlington	50	143	460	23	242	45%*	55%
With Ferric Chloride Addition							
Sarnia	50	68	1200	60	179	20%	18%
Windsor	50	165	600	30	191	17%	18%
Burlington	50	143	460	23	280	15%*	18%
With Ferric Chloride and Polymer Addition							
Sarnia	50	68	1200	60	167	8%	9%
Windsor	50	165	600	30	202	10%	9%
Burlington	50	143	460	23	308	10%*	9%

Note: For comparison purposes, the results shown in this table were interpolated to bring actual detention time to a constant figure for all three settling tanks.

* Extrapolated results

PREDICTION

The settling model is based on studies carried out at the treatment plants in Sarnia and Windsor and at the pilot plant in CCIW, Burlington, with and without the addition of chemicals to domestic wastewaters. It is thought to be applicable to other plants treating similar wastewaters. However, further verification is required. The performance of primary clarifiers can be predicted from the following relationship:

$$S = \frac{S_o A}{(t_g)^n + A} \qquad \text{Settling Model}$$

where

S	=	Suspended solids remaining in effluent of settling tank, mg/l
S_o	=	Suspended solids in infleunt of settling tank, mg/l
t_g	=	Actual mean detention time, minutes, obtained from tracer studies
n, A	=	Constants of wastewater characteristics developed from settling column tests

The value of the constants 'n' and 'A' for <u>domestic</u> wastewater, with and without chemical addition are:

Chemical Addition	n	A
No chemical	0.64	15.0
Ferric chloride (or alum)	0.90	7.5
Ferric chloride (or alum) plus Polymer	0.93	3.5

For other suspensions the constants must be determined from settling tests.

Limitations on the use of the settling model for domestic wastewater:

- t_i, time interval for initial indication of tracer in effluent, should not be less than 15 minutes with chemical, and 30

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minutes without chemical additions. This time is needed for flocculation, development and settling of flocs, particularly when chemical treatment is practised. For t_i values less than the above, calculate suspended solids remaining in effluent directly from S-curve and C-curve.
- overflow rates should be less than 2000 gpd/ft^2 (100 m^3/m^2/day)

Operational constraints at the plants did not allow work at overflow rates in excess of 2200 gpd/ft^2 (110 m^3/m^2/day). Gray (1974) reported excessive carry-over of solids at the Sarnia plant at an overflow rate of 2090 gpd/ft^2 (104 m^3/m^2/day) contrary to experience in this study. Gray did not use polymer. At 2200 gpd/ft^2 (110 m^3/m^2/day), removal of suspended solids was 34, 72 and 87% without, with ferric chloride, and with ferric chloride plus polymer additions, respectively. However, an upper limit on overflow rate undoubtedly exists. Further work is required to establish these limits.

TENTATIVE DESIGN GUIDELINES

The following tentative guidelines are proposed for the design of primary clarifiers (horizontal flow clarifiers) treating domestic sewage:

Design flow rate: Maximum daily flow (normally 1.5 - 2.0 x average daily flow)

Actual <u>mean</u> detention time (t_g):
         30 minutes with chemical addition (ferric chloride plus polymer)
         45 minutes without chemical addition

Minimum detention time (t_i):
(t_i time interval for initial indication of tracer)
         10 minutes with chemical addition (ferric chloride plus polymer)
         20 minutes without chemical addition

Overflow rate: less than 2000 gpd/ft^2 under peak flow conditions.

Velocity: less than 8 ft/minute (40 mm/sec) under peak
                          flow conditions.

Removal efficiency: about 70% with ferric chloride addition
                    about 85% with ferric chloride plus polymer addition
                    about 40-50% without chemical addition

 Laboratory results and limited plant results on alum indicate
that the above design guidelines may also be applicable for alum addition.

 It is beyond the scope of this paper to make recommendations
regarding the geometric design of settling tanks (shape, length/width/
depth or diameter/depth ratios, types and positions of baffles and
weirs, etc.). The problem remains, therefore, how to choose actual
tank dimensions to achieve a required actual detention time.

 The use of these tentative guidelines will result in smaller
settling tanks than would be required by current guidelines, discussed
previously. The extent of it will depend on the hydraulic efficiency,
e, which was previously defined as the ratio of actual mean detention
time to theoretical detention time. The hydraulic efficiency of tanks
studied under varying flow rates ranged from 0.30 (circular tank at
Windsor) to 0.78 (rectangular tank at Windsor) to 0.78 (rectangular tank
at Sarnia).

COST IMPLICATIONS

 Chemical treatment is coming into widespread use, primarily
for removing phosphorus from wastewaters. This study revealed that
such chemical addition has considerable secondary benefits. Addition
of chemicals causes suspensions to settle faster and, consequently,
significantly smaller settling tanks are required than with no chemical
addition. Furthermore, with chemical addition, more reliable removal
efficiency of suspended solids and BOD, as well as phosphorus, is
achieved.

 If the findings of this study are found to be applicable
generally, chemical treatment will result in considerable savings in
capital costs which, in turn, will offset part or all of the operating
cost of chemical addition. For existing plants, chemical addition
will substantially increase the capacity of the settling tanks, avoid-

ing or delaying the need for expansion. For example, as a result of this study, the Sarnia treatment plant has been re-rated at about 14 MGD capacity (previous rating 8 MGD). In some cases effluent quality may be good enough to be discharged into receiving waters without any secondary treatment.

The beneficial effects of chemical treatment are not limited to primary treatment only. If a treatment plant has both primary and secondary systems, the BOD and SS loads to the secondary system will be reduced considerably because of the higher rates of BOD and SS reduction at the primary stage. (BOD removal increases from 20-30% range to 60-70% range; SS removal from 40-50% range to 80-90% range.) Consequently, the capacity of the secondary system will also be increased.

Chemical treatment, however, results in greater quantities of sludge than are produced in a conventional primary treatment plant, with resulting increases in the cost of handling excess sludge.

CONCLUSIONS AND RECOMMENDATIONS

It is concluded that:

1. Addition of chemicals (ferric chloride or alum) increases settling rate of physical-chemical flocs and improves effluent quality significantly. Addition of polymer in combination with ferric chloride or alum further increases settling rates and effluent quality.

2. The clarification efficiency of a settling tank treating flocculent suspensions depends primarily on the actual mean detention time of the suspension flowing through the tank. Overflow rate up to about 2000 gpd/ft^2 (100 m^3/m^2/day) does not have a significant effect on effluent quality, particularly when chemicals are added.

3. Settling column and tracer tests are useful tools to evaluate the performance of settling tanks.

4. The costs incurred for chemical treatment of sewage for phosphorus reduction will be partly offset through reduced capital costs for the primary and secondary treatment plant.

In certain circumstances chemical treatment in a primary treatment plant may eliminate the need for secondary treatment.

It is recommended that:

1. Similar studies be carried out at other treatment plants and to include the following:

   - horizontal and upflow clarifiers;
   - the use of lime, alum and polymers;
   - hydraulic efficiency of clarifiers; and,
   - functional design of tanks.

2. Current design guidelines for primary clarifiers be reviewed based on the findings of this and future studies.

REFERENCES

ASCE and WPCF, "Sewage Treatment Plant Design", Joint Committee Report, Manual of Practice No. 8, (1959) reprinted 1967.

Burns and Roe Inc., "Process Design Manual for Suspended Solids Removal" EPA Technology Transfer, Washington, Oct. 1971.

Camp, T.R., "Sedimentation and the Design of Settling Tanks", Trans. ASCE, 111, 895 (1946).

Clements, M.S., 1966, "Velocity Variations in Rectangular Sedimentation Tanks", Proc. Inst. Civil Eng. 34, 171.

Clements, M.S., and Khattab, A.F., 1968, "Research into Time Ratio in Radial Flow Sedimentation Tanks", Proc. Inst. Civil Eng., 40, 471.

Fielder, R.A., and Fitch, E.G., 1959, "Appraising Basin Performance From Dye Test Results", Jour. Water Pollution Control Federation, 31, 9, 1016.

Gray, I.M., "Phosphorus Removal Study at the Sarnia WPCP", Report No. 14, Canada-Ontario Agreement on Great Lakes Water Quality, Ottawa and Toronto, 1974, 22 pp.

Heinke, G.W., "Design and Performance Criteria for Settling Tanks for the Removal of Physical-Chemical Flocs", Res. Report No. 10, Canada-Ontario Agreement on Great Lakes Water Quality, Ottawa and Toronto, 1973, 82 pp.

Heinke, G.W., Qazi, M.A., and Tay, A., "Design and Performance Criteria for Settling Tanks for the Removal of Physical-Chemical Flocs", Final Report, March 1975, 106 pp. University of Toronto, Department of Civil Eng. To be published under Canada-Ontario Agreement on Great Lakes Water Quality.

Kalbskopf, K.H., "European Practices in Sedimentation" in Water Quality Improvement by Physical and Chemical Processes, edited by E.F. Gloyna and W.W. Eckenfelder Jr., University of Texas Press, Austin and London (1970), p. 92.

MacLaren, James F. Ltd., "To establish viable methods of maintaining waste treatment facility efficiencies with reference to flow variations", Res. Report No. 11, Canada-Ontario Agreement on Great Lakes Water Quality, Ottawa and Toronto, 1973.

Price, G.A., 1974, "An Extension of the Time Ratio Theory of Sedimentation Tanks", Proc. Inst. Civil Eng., 50, 191.

Roy F. Weston, Inc., "Process Design Manual for Upgrading Existing Wastewater Treatment Plants", EPA Technology Transfer, Washington, Oct. 1971.

Villemont, J.R., Rohlich, G.A., and Wallace, A.T., 1967, "Hydraulic and Removal Efficiencies in Sedimentation Basins". In "Advances in Water Pollution Research", Proc. Third Intl. Conf. Water Pollution Res., Munich, Germany, Water Pollution Control Fed., Washington, D.C., 2, 381.

Voshel, D. and Sak, J.G., "Effect of Primary Effluent Suspended Solids and BOD on Activated Sludge Production", Jour. Wat. Poll. Cont. Fed., 40, R203 (1968).

UPGRADING LAGOON EFFLUENTS

by

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Throughout the Province of Ontario and especially in south-western Ontario, lagoons or stabilization ponds serve as the wastewater treatment facility for a large number of smaller communities.

In many cases, in order to improve the receiving stream water quality, continuous flow lagoons have been converted to seasonal discharge systems, effectively reducing their treatment capacity because of the required 180 day detention period to facilitate twice per year discharge.

The objective of this study was to determine if the effluent quality from a standard lagoon or stabilization pond could be significantly improved by the addition of post treatment units at the effluent end of the system, thereby eliminating the need for seasonal discharge. A standard lagoon was defined as a continuous discharge system having a detention time of approximately 120 days with an organic loading of 20 lb BOD/acre/day. The expected effluent quality from such a system was 20 mg/l BOD and SS, with 5 mg/l ammonia and phosphorus.

From information available within the Ontario Ministry of the Environment, a suitable site for such a program was determined to be Strathroy, Ontario, a town of 7,000 people, location in Middlesex County in southwestern Ontario.

Because of the expected change in effluent quality associated with the change in temperature from winter to summer, the program was scheduled to begin in the fall and carry on for one full year. Start up and delivery problems delayed the project until January 1975 resulting in a comprehensive 9 month study with 3 months of preliminary data from October to December 1974.

SUSPENDED SOLIDS REMOVAL

It was obvious that the major pollutional components in the lagoon effluent were associated with the suspended solids being discharged. Thus three types of suspended solids removal systems were investigated.

1. Microfloc pilot plant

This system incorporated flash mixing, flocculation, sedimentation and final effluent polishing by multimedia filtration. The flow rate was established at 10 gpm throughout the period of study. Several runs were made without chemical additions but during most of the runs the addition of alum, ferric chloride and/or polymer was practiced to enhance suspended solids removed by flocculation and sedimentation.

2. Flotation

A small pilot dissolved air flotation unit was employed in an attempt to float the algae to the surface and remove it by surface scum removal. The addition of coagulants was also attempted to improve the solids capture.

3. Microstraining

A pilot microstrainer using a 23 micron (Mark 0) stainless steel cloth and handling flow rates up to 30 gpm was used for suspended solids removal. The addition of coagulants was attempted in order to improve the solids capture and the distribution of product and reject waters. The major portion of this phase of the study was carried out during the summer when lagoon effluent suspended solids levels were high.

REMOVAL OF SOLUBLE COMPONENTS

Work was also done to eliminate the soluble pollutional component of the effluents, namely, ammonia nitrogen, phosphorous and dissolved BOD. Chemical addition was used for phosphorus removal with two alternate biological systems used to reduce the ammonia and soluble BOD levels.

DISCUSSION OF RESULTS

General

As seen from Figure 1, the organic loading on the Strathroy lagoon varied from a low of 16 lb BOD per acre per day during April and July to a high of 27 lb BOD per acre per day during May, with an average loading of 21 lb BOD per acre per day based on the total raw sewage flow being diverted to the north cell (38 acres). Raw sewage samples were taken every three days using a flow proportional 24 hour composite sampler. Flow measurement was done using a V notch weir and an integrating recorder which was read daily. The flow was checked daily against calibrated pump running times. The two techniques of flow measurement provided almost identical numbers.

The detention time of the lagoon was calculated based on the recorded flow and the verified lagoon volume to be as shown in Figure 1. The minimum detention time of 86 days occurred during the spring runoff period in March and April with a maximum detention of 110 days during July. The average detention time was 100 days which is slightly lower than the accepted standard of 120 days.

In general, it can be said that the Strathroy lagoon system was operating at its design organic load and slightly above its hydraulic load. It was felt that this system adequately represented the "standard" waste treatment lagoon.

The effluent quality from the lagoon is shown in Figures 2, 4, 6 and 7.

As was expected, the seasonal variation in BOD and SS was high with the BOD ranging from a maximum monthly average of 21 mg/l to a minimum of 12 mg/l and the suspended solids ranging from 12 mg/l to 47 mg/l. Figure 6 shows high winter phosphorus levels of 5.2 mg/l with a minimum value of 2.6 mg/l in May. The effluent ammonia values, Figure 7, show a similar seasonal variation from 10.5 mg/l in March to a minimum of 4 mg/l in June.

An unexpected result is shown in Figure 2 where the soluble BOD shows only a slight seasonal variation with a range of 3.5 to 7 mg/l dissolved BOD.
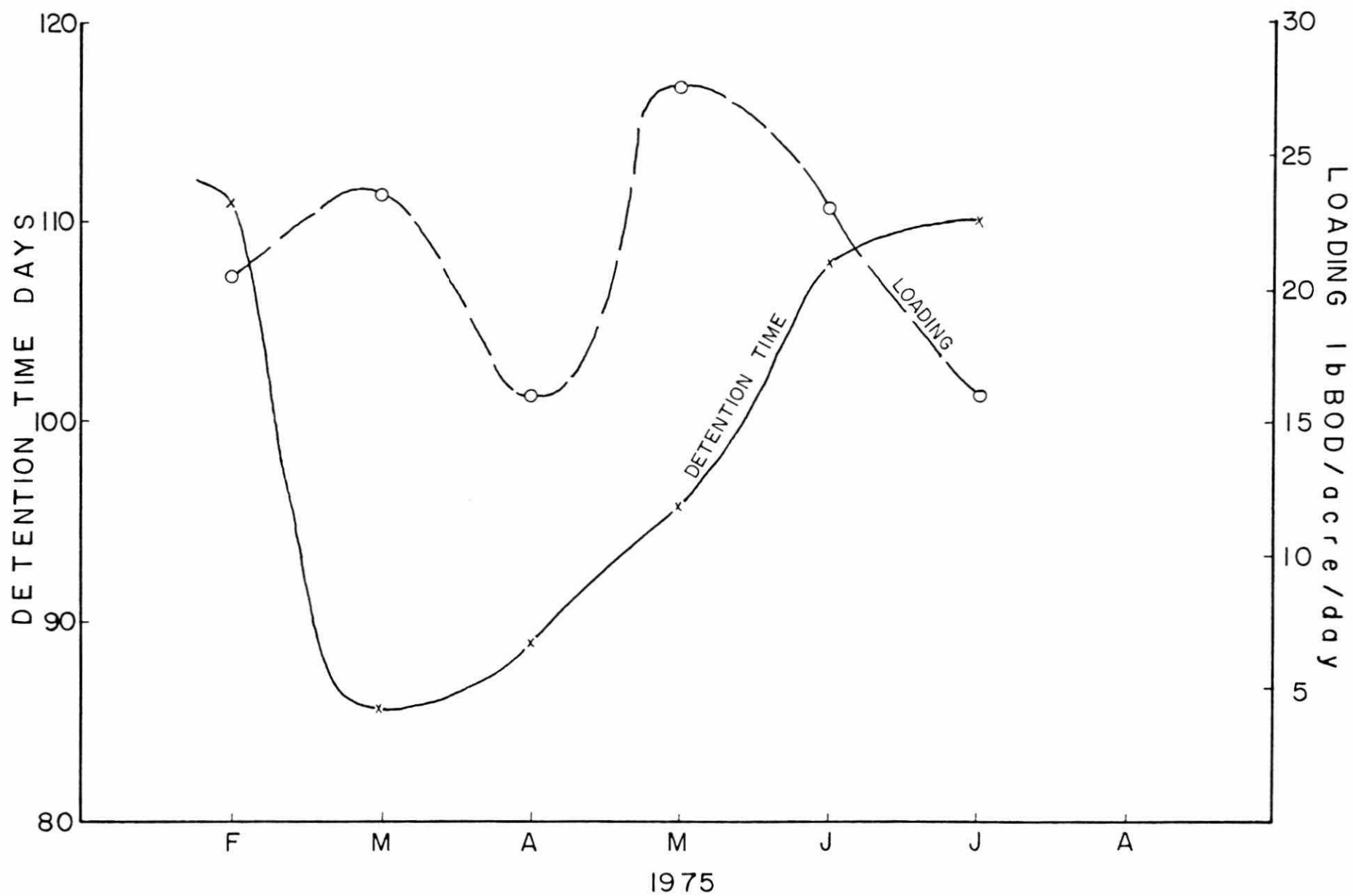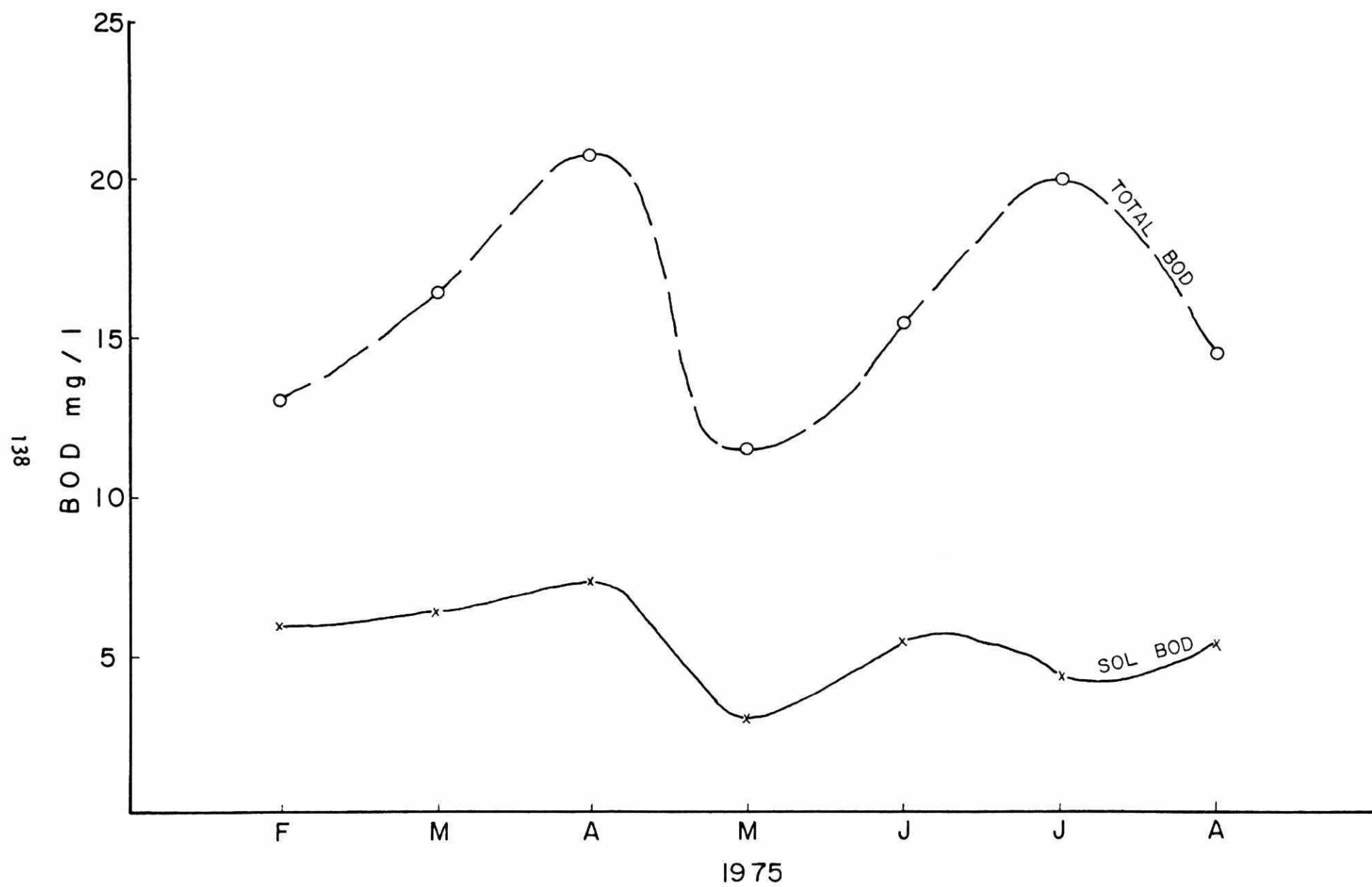
FIGURE 1. LAGOON LOADING AND DETENTION

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FIGURE 2.   LAGOON EFFLUENT TOTAL AND SOLUBLE BOD.

It had been expected that high seasonal variations in soluble BOD would be experienced because of the low winter operating temperature and the long period of ice cover (January to March).

DO levels in the lagoon effluent were always high, 6 to 12 mg/l, as long as the lagoon was ice free. During the winter periods, low DO levels (less that 1.0 mg/l) and sulphide odours were evident in the effluent stream. The effluent withdrawal point is located approximately one foot from the bottom of the five foot deep lagoon.

Suspended Solids Removal

A. Microfloc Pilot Plant.
The reason for using the Microfloc pilot plant was twofold. The unit is capable of removing suspended solids and at the same time, because of the addition of prime coagulants to enhance the solids removal, phosphorus removal will be effected.

Standard jar tests were done on the lagoon effluent several times per month. The response of the effluent to chemical addition tended to vary rather markedly throughout the period of the study. In fact, the optimim chemical dosage could vary up to 40% higher or lower within a two or three day period. It was felt that the large changes in coagulation response were due to the changes in algae concentrations and types in the lagoon.

Because of the rapidly changing coagulation characteristics of the effluent, it was at times difficult to control the microfloc system and thus optimum performance was not always maintained. The pilot unit was operated continuously (24 hours per day) throughout most of the study. Automatic pressure recorders provided the run time and backwash frequency data.

As seen in Figure 3, in spite of the difficulty described above, a reasonably consistent effluent BOD was maintained from the Microfloc system. The BOD values ranged from a low of 6 mg/l to a high of 9 mg/l. By comparing the soluble BOD shown in Figure 2 and the microfloc effluent BOD shown in Figure 3, it is obvious that the microfloc type system, when properly operated, can approach the soluble BOD level quite consistently.

This is confirmed by the suspended solid removal capability shown in Figure 4. In spite of high feed suspended solids levels in
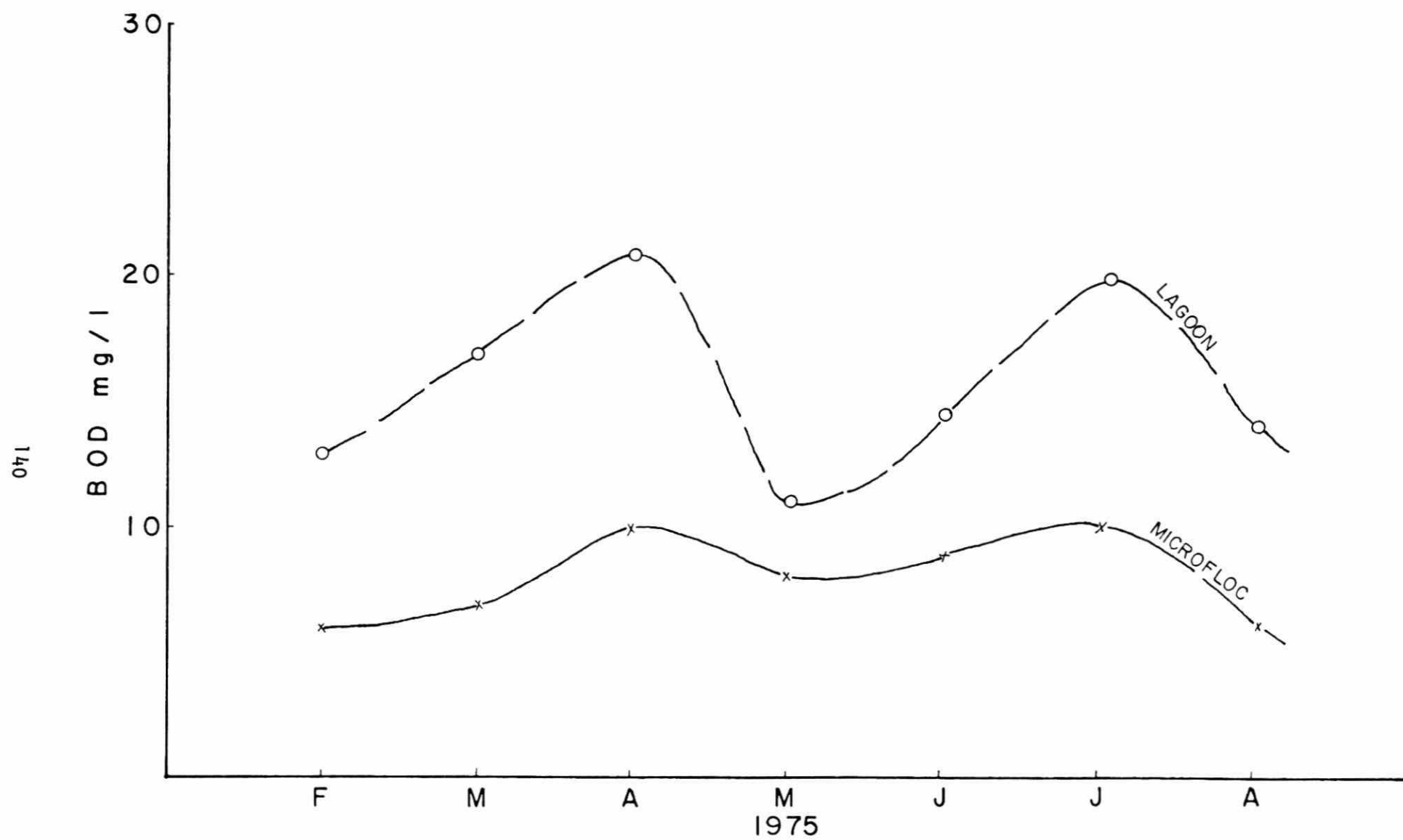
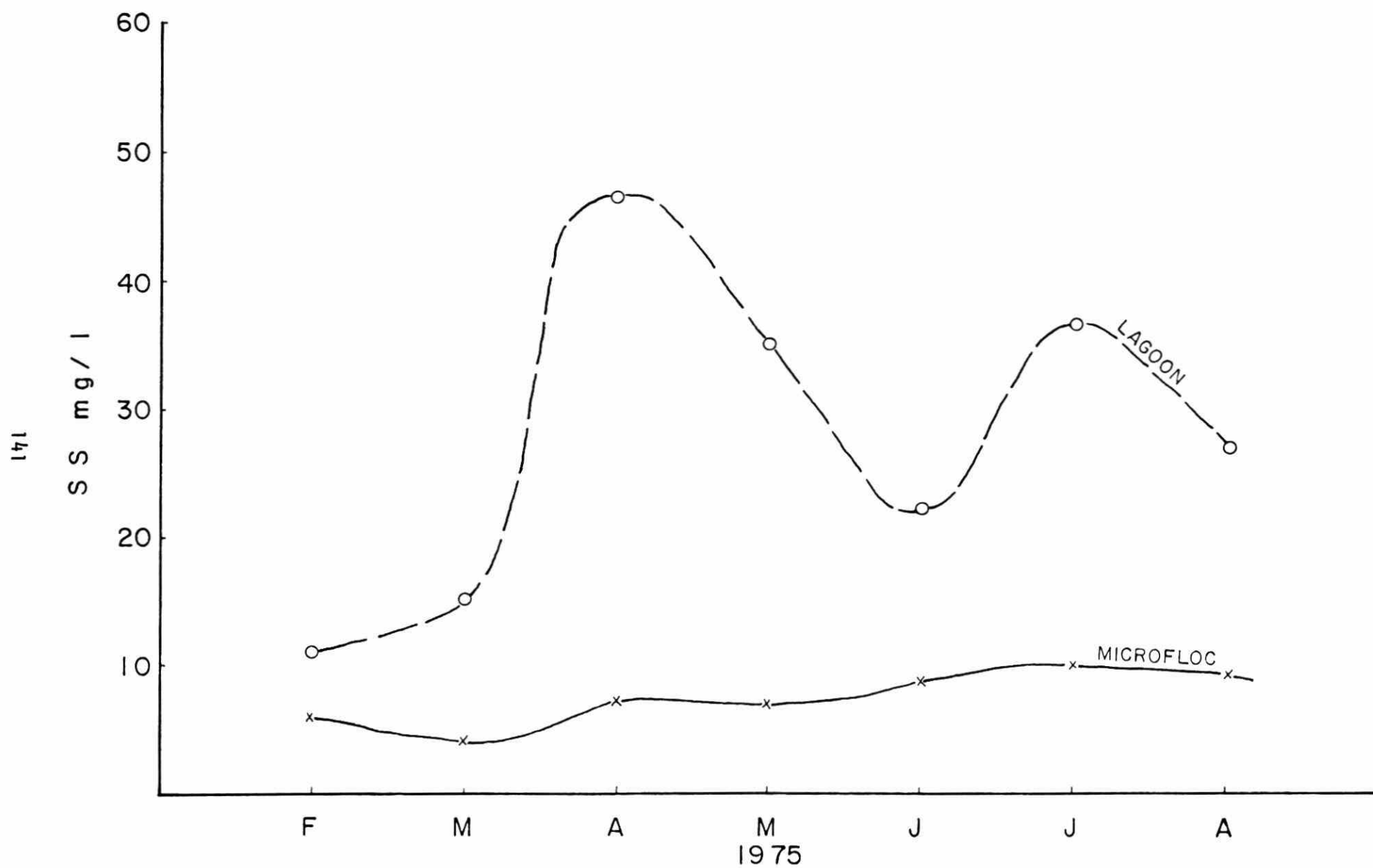FIGURE 3. MICROFLOC PERFORMANCE BOD REMOVAL.

FIGURE 4. MICROFLOC PERFORMANCE SUSPENDED SOLIDS REMOVAL

April, the unit effluent remained less than 10 mg/l SS at all times. Phosphorus removal to less than 1 mg/l total P was also consistently achieved, (Figure 7).

Operation of the multimedia filter was assessed on a separate unit basis with both filter feed and effluent being monitored. Figure 5 clearly demonstrates that when running the filter at a constant rate of 5 gpm/ft^2 and a fixed terminal head loss of 5 ft of H_2O, a definite relationship exists between the filter feed suspended solids level and the duration of the filter run. As the feed suspended solids approach 30 mg/l the filter run time drops to less than two hours. Thus it is obvious that coagulation and sedimentation is required prior to filtration to ensure reasonable filter run times.

Figure 6 shows a poorly defined relationship between influent and effluent suspended solids with the effluent solids being approximately 40% of the filter feed solids level. The data scatter is rather wide and thus a firm relationship cannot be established. It is likely that a well coagulated filter feed would more clearly define this relationship.

It can be concluded that a coagulation sedimentation, filtration system can effectively convert a conventional lagoon effluent into a high quality effluent with BOD and SS levels below 10 mg/l and less than 1 mg/l total phosphorus. No ammonia conversion or reduction can be expected from the above physical - chemical system. Effluent ammonia levels remain high at 4 to 10 mg/l.

B. Microstrainer. A Crane pilot microstrainer was used to treat lagoon effluents for a three month period from July to September of 1975. Initial results without the addition of prime coagulants resulted in from 20% to 50% solids removal and a corresponding 10% to 20% BOD reduction. Actual microstrainer effluent values were from 10 to 20 mg/l BOD and SS. In an attempt to improve the solids removal efficiency of the unit, coagulation of the feed with alum or ferric chloride and polyelectrolyte was tried. An intermittent improvement in solids capture was noted with some runs showing 70% SS removals to 16 mg/l and many runs showing an increase in microstrainer effluent solids or a negative percent removal (to 100 mg/l).
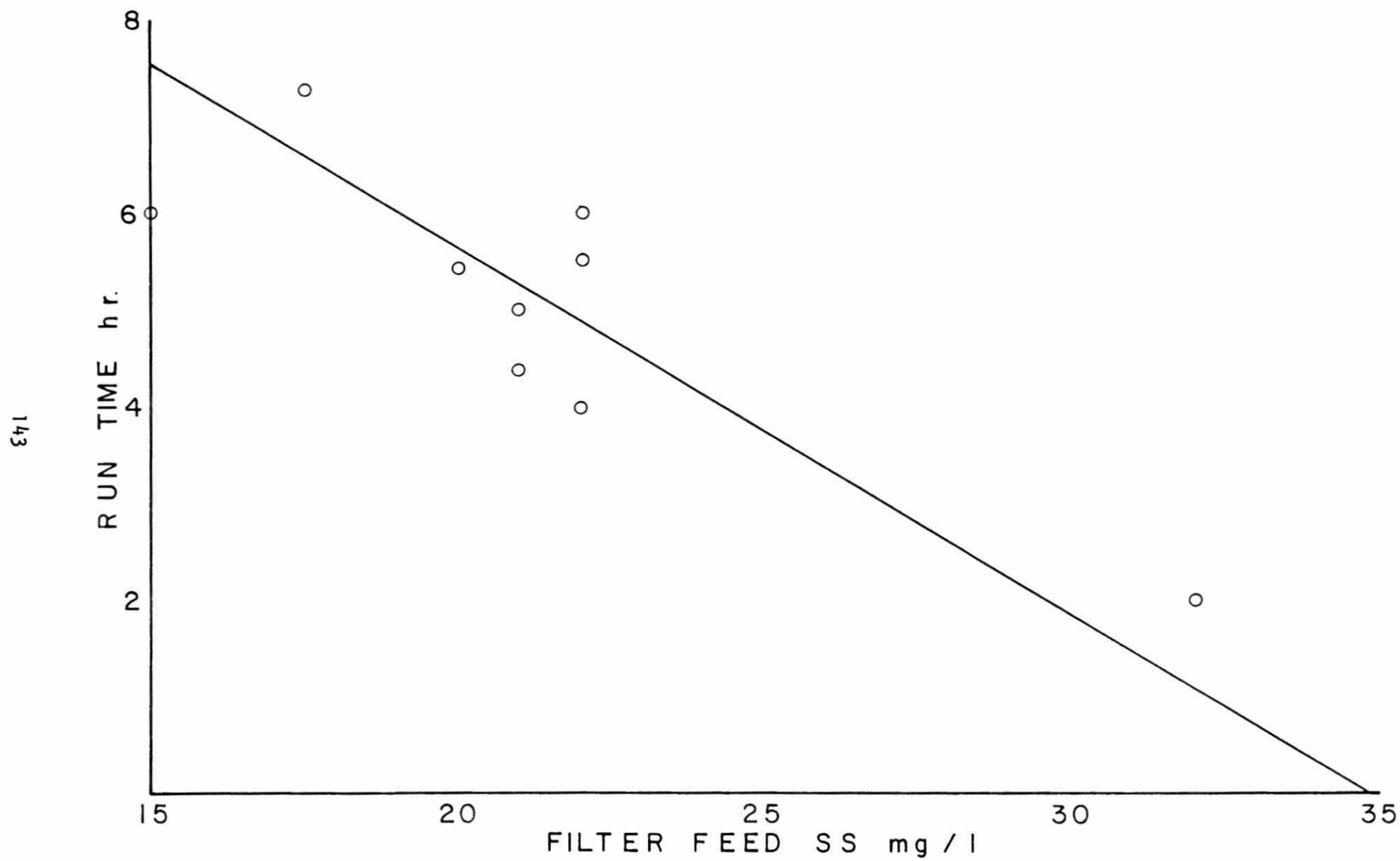
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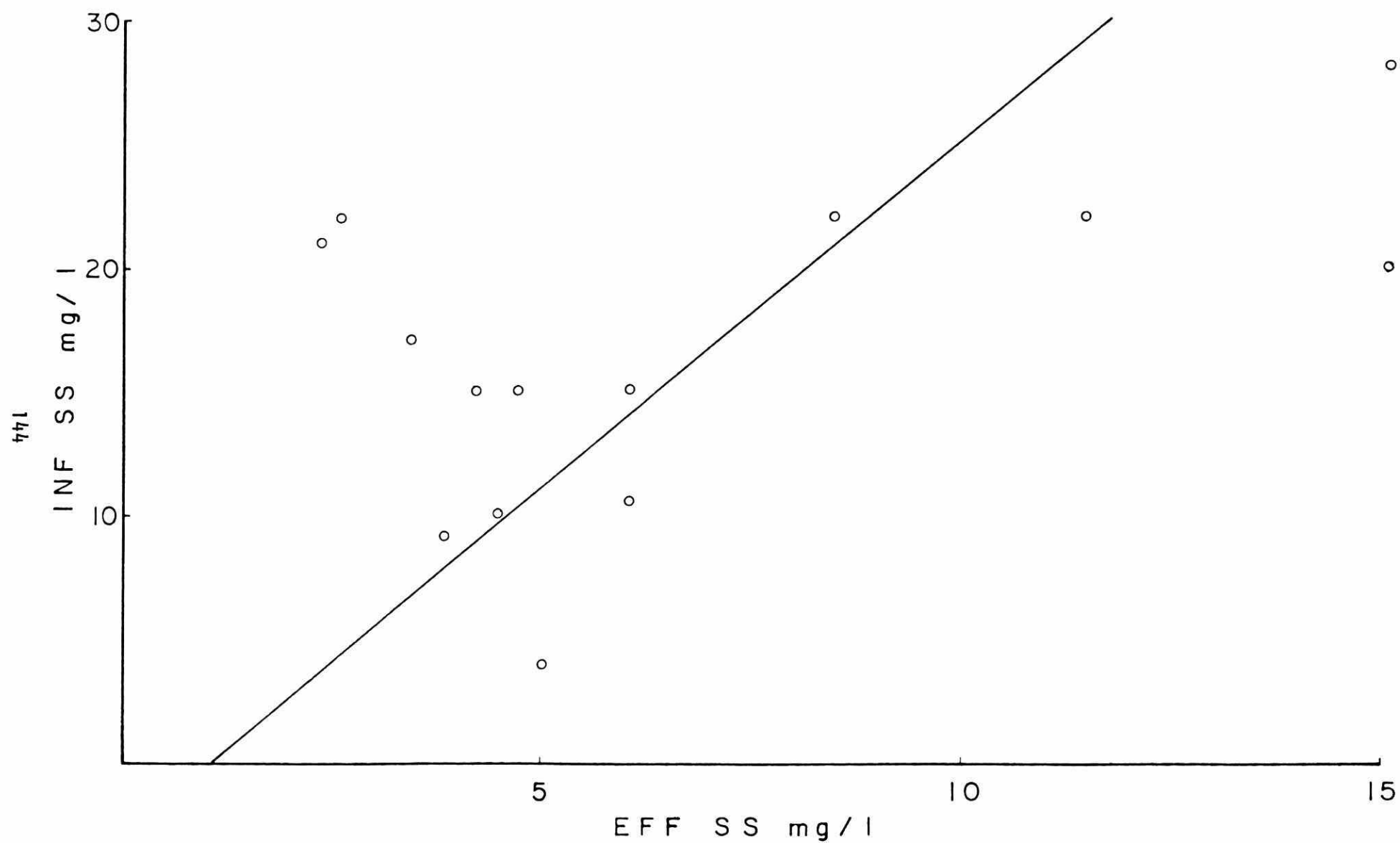FIGURE 5. MULTIMEDIA FILTER RUN TIME.

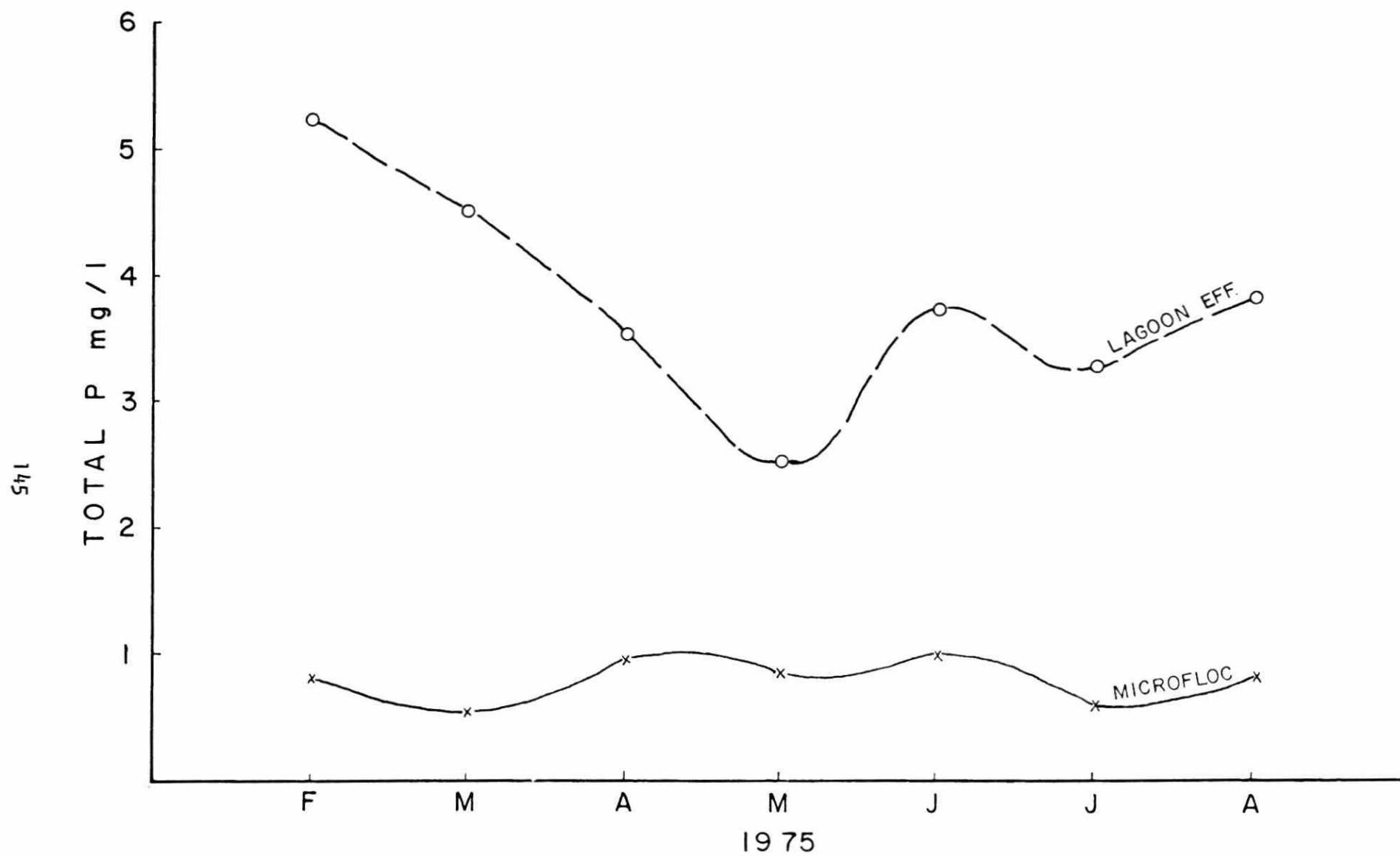FIGURE 6.   MULTIMEDIA FILTER PERFORMANCE SUSPENDED SOLIDS REMOVAL.

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FIGURE 7. MICROFLOC PERFORMANCE PHOSPHORUS REMOVAL.

BOD removal was consistently positive, varying from 25 to 70%. Actual effluent values ranged from 6 to 16 mg/l.

In both operating modes the microstrainer was not capable of producing the high quality effluent desired in this study. Its erratic performance and frequent need of batchwise cleaning with chlorine solution to improve its hydraulic capacity make it an unsuitable system for polishing of lagoon effluents.

C. Dissolved Air Flotation. The flotation unit was operated from March to July, 1975, and consistent 20 to 30% BOD removal down to 8 to 12 mg/l BOD was maintained. The suspended solids removal, however, was erratic varying from negative removals to 60% removal. During the critical summer period, flotation effluent solid levels were always high at 15 to 30 mg/l.

The effluent phosphorus levels were 1.0 to 2.0 mg/l, being reduced by the prime coagulants used in the flotation system.

This unit was capable of continuous operation, producing an intermediate level of suspended solids, BOD and phosphorus removal. It does not produce the required high quality effluent desired.

Reduction of Soluble Components

Conversion of ammonia nitrogen to nitrate nitrogen is desirable in attaining a minimum oxygen demand in the effluent. Each pound of ammonia nitrogen discharged exerts 4.5 lb of oxygen demand on the receiving stream. The objective of ammonia conversion was to reduce the ammonia level from the 4 to 10 mg/l found in the lagoon effluent to less than 1 mg/l. Figure 8 shows that the effluent ammonia level varies seasonally from a minimum value of 4 mg/l in June to a maximum value of 10 mg/l in March. This variation is likely due to the reduced atmospheric losses of ammonia because of the winter ice cover. At no time was there more than 1 mg/l nitrate nitrogen in the lagoon effluent in spite of the consistently high raw sewage ammonia levels of 20 - 30 mg/l, indicating negligible biological ammonia conversion to nitrate. Thus there was an unexplained loss of from 10 - 25 mg/l of nitrogen across the lagoon system.

Direct losses of ammonia nitrogen to the atmosphere are a distinct possibility since the lagoon pH levels frequently rise to 10 or

more. At this pH level, ammonia exists primarily in the more volatile non-ionic form NH_3.

R B C Pilot Plants

1. Bio-Surf System. Pilot work centered around the biological conversion of ammonia to nitrate on rotating biological contactors (RBC). Two types of contacts were used. The first was a RBC unit supplied by the Autotrol Corp. known as the Bio-Surf system. This unit was started in January 1975 at a low flow rate of 0.5 gpd/ft^2 of surface available for biological growth. Lagoon effluent was used as feed. As can be seen from Figure 8, a three month startup period was required before adequate nitrification was begun. This extended startup period was likely due to the very low (4° C) operating temperatures experienced throughout the startup period. Flows were maintained at 0.5 gpd/ft^2 throughout the startup period. From mid April to the end of May the flow was doubled on four occasions bringing the final flow to 8 gpd/ft^2. At this final flow level a slight deterioration in Bio-Surf effluent ammonia levels was noted with an increase from 0.3 mg/l to 0.9 mg/l. This flow was maintained throughout the summer with a consistently good ammonia removal being attained at an operating temperature of 18 to 23° C.

Table 1 shows the corresponding BOD, Sol. BOD and SS removal experienced during summer operations. Approximately 20% BOD and SS reduction was noted. Virtually no soluble BOD removal was achieved. Attempts were made to settle the Bio-Surf effluent with no success. Unsettled solids levels were approximately 24 mg/l with no decrease being noted even when held under quiescent conditions for two hours.

2. Euro-Matic Biodrum. For a two month period during the summer a RBC pilot unit as supplied by Prospereau Consultation Ltee was used for ammonia conversion. The results of this study are shown in Figure 9.

The startup period was significantly shorter than the cold weather startup of the other RBC pilot unit, with full nitrification being achieved within three weeks.

The unit was started at a low loading of 3.5 gpd/ft^2 and gradually increased to 9 gpd/ft^2, at which level it performed well with no evidence of a deteriorating effluent quality.
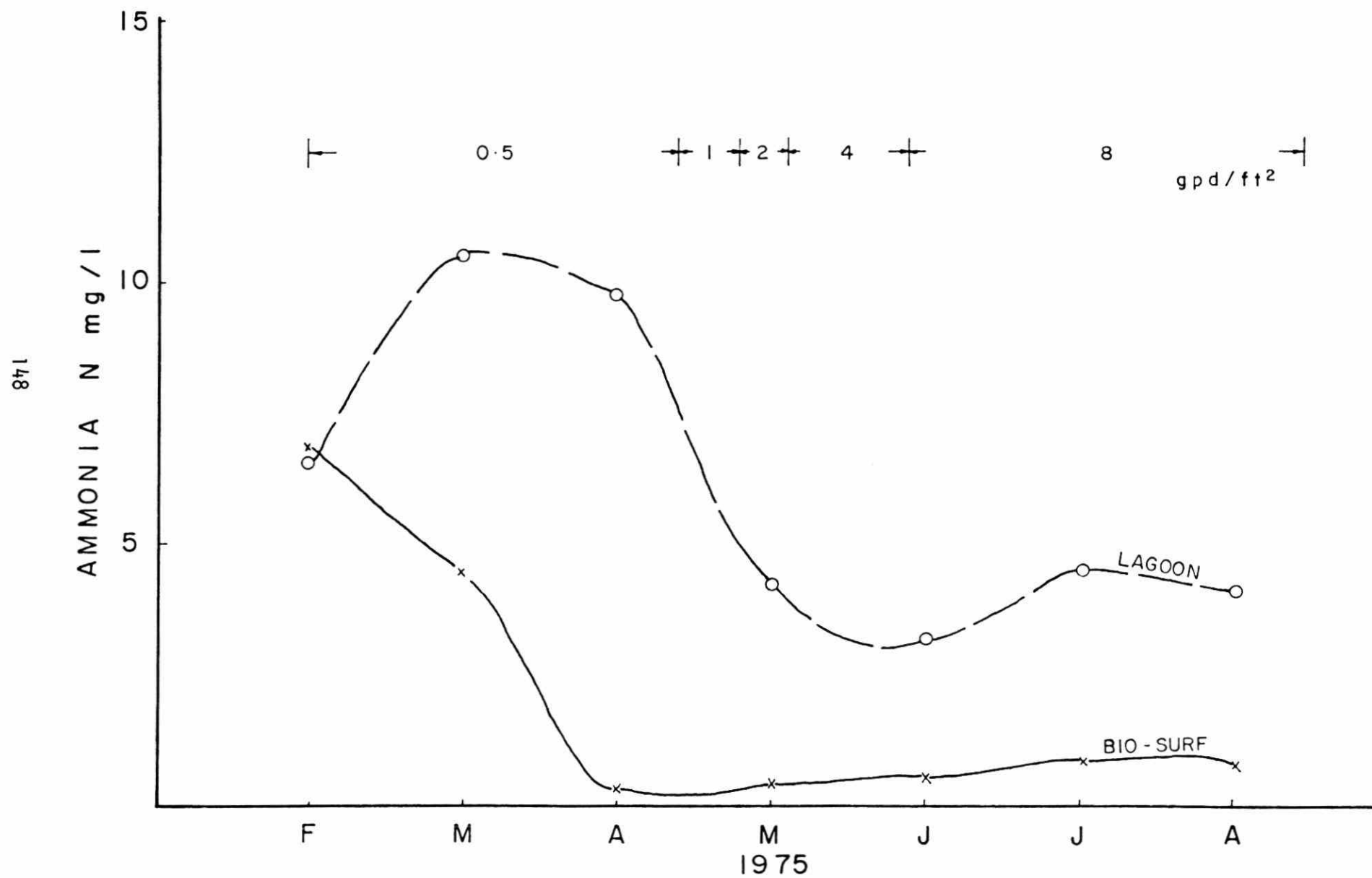
FIGURE 8.   BIO-SURF PERFORMANCE AMMONIA CONVERSION
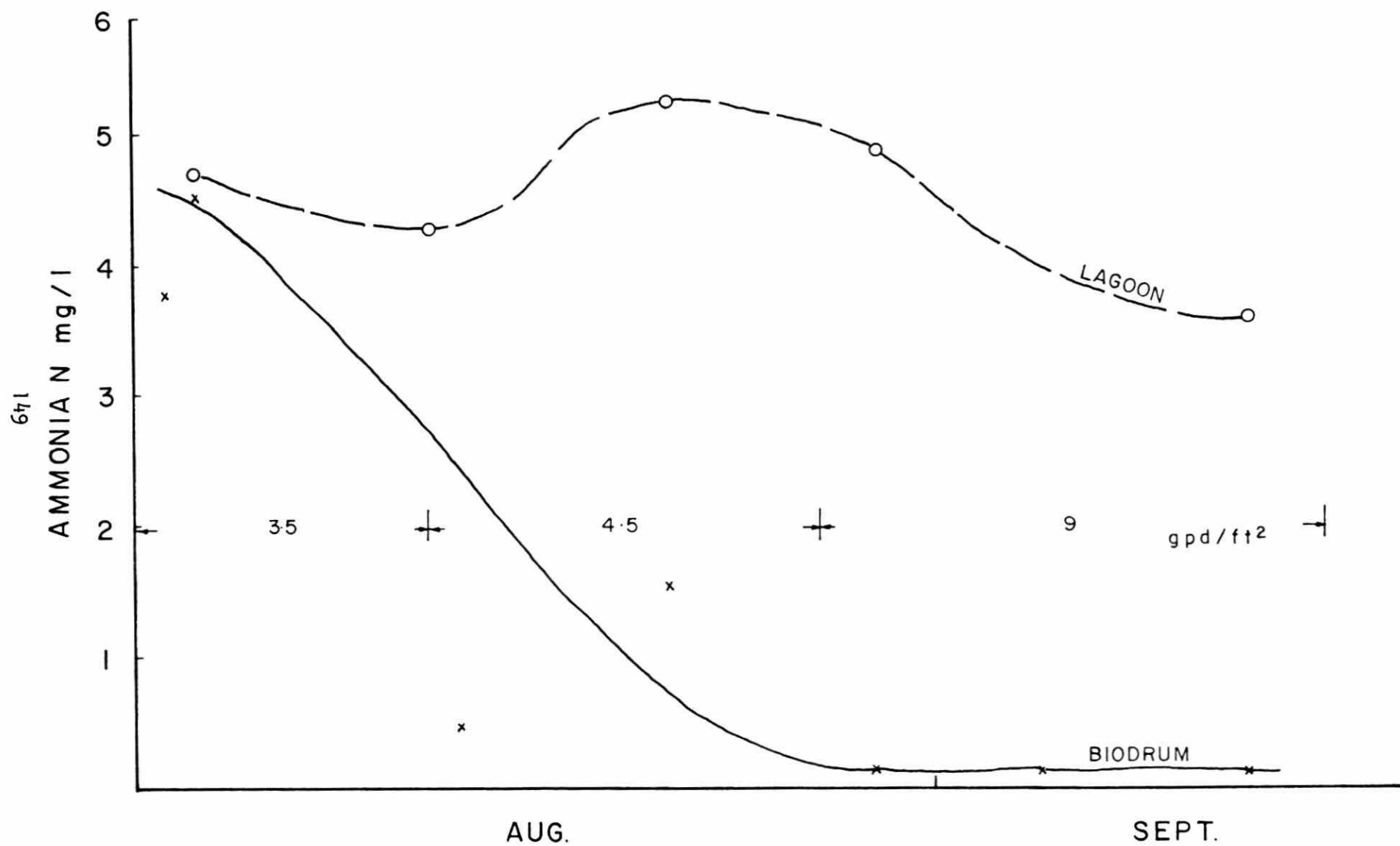
FIGURE 9.   BIODRUM PERFORMANCE AMMONIA CONVERSION

TABLE 1. BIO-SURF OPERATION EFFLUENT VALUES AND
REMOVAL EFFICIENCIES.

Month	BOD				SS	
	Total		Sol			
	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal
May	11	8	5	0	25	28
June	13	19	6	0	20	9
July	16	20	5	0	28	22
Aug.	14	0	4	30	16	43

TABLE 2. BIO DRUM OPERATION EFFLUENT VALUES AND
REMOVAL EFFICIENCIES FOR AUGUST AND
SEPTEMBER 1975.

	Effluent mg/l	% Removal
BOD	15	0
Soluble BOD	5	15
Total PO_4 as P	3.4	5
SS	26	7
NH_3 as N	0.1	98

With respect to BOD, SS and phosphorus removal, the unit provided virtually no change in these parameters as seen in Table 2.

From both these studies, it can be concluded that lagoon effluent ammonia levels can be effectively reduced to less than 1.0 mg/l using RBC equipment operating at high hydraulic loading rates (8 to 9 gpd/ft^2) during the critical summer period.

Startup times of approximately three weeks can be expected.

RECOMMENDED SYSTEM AND COSTS

Based on the data collected in a nine month study at the Strathroy lagoon, the following appears to be an effective post treatment system for lagoons.

Figure 10 shows a line drawing and basic design paramaters of an ammonia conversion, solids removal and phosphorous removal post treatment system.

It incorporates a rotating biological contactor, flash mixing, sedimentation and filtration.

The expected effluent quality is 7 mg/l BOD, 7 mg/l SS and 1.0 mg/l ammonia.

Because of the volume of the lagoon, it will be possible to run such a post treatment system at a fixed rate for extended periods of time. This should aid in minimizing the requirement to overdesign to handle maximum daily hydraulic loads and should make operation somewhat easier. It will be necessary to provide considerable manpower for operation because of the need to continuously monitor the effectiveness of the chemical system, as day to day variations can be expected.

If winter operation is required, the system should be totally enclosed as the liquid temperatures will be 4° C throughout the ice covered period. Ammonia conversion throughout the winter is not practical with this biological conversion system. It should only be run during the ice free period.

Consideration should be given to the overall impact of such an effluent on the receiving stream. The total measurable oxygen demand will be approximately 7.0 mg/l from the BOD and 4.5 mg/l from the ammonia, or 11.5 mg/l O_2 in total. During the summer period when the effluent DO

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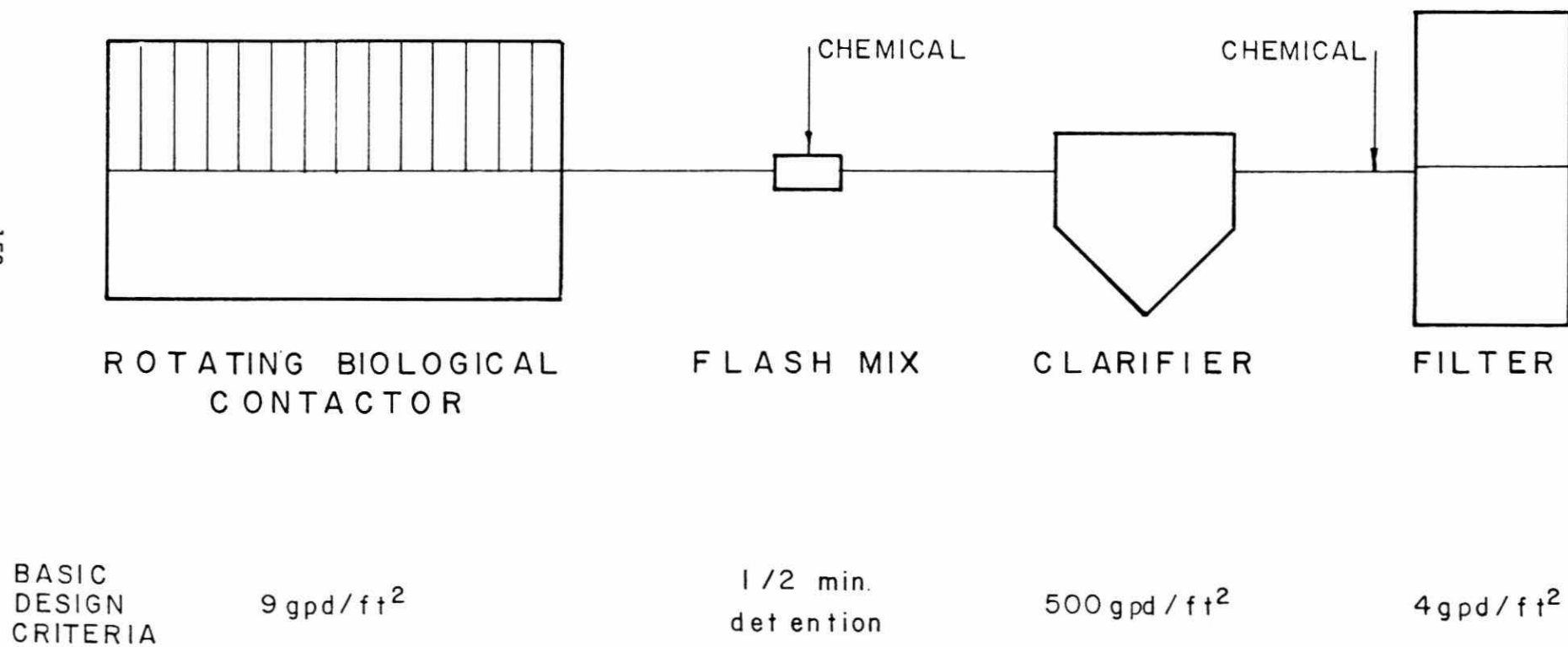| BASIC
DESIGN
CRITERIA | 9 gpd/ft^2 | 1/2 min.
detention | 500 gpd/ft^2 | 4 gpd/ft^2 |

FIGURE 10.   PROPOSED SYSTEM

levels will be 10 to 12 mg/l the net effect on the stream should be neutral or it might have a positive effect when the oxygen input exceeds the oxygen demand.

Consideration might be given to providing a short post aeration facility to supplement the oxygen levels when they fall below saturation during extended cloudy periods or periods of ice cover.

The costs associated with this system are shown in Table 3.

TABLE 3.   COSTS

	Operating ¢/1000 gal	Capital $	Operating ¢/1000 gal	Capital $
Rotating Biological Contactor *	3	80,000	2.5	220,000
Clarification **	9	650,000	7.5	1,400,000
Filtration **	9	300,000	6	550,000

*   From Napier-Reid Ltd. Toronto.

** Based on ENR construction Index of 2200. <u>1975</u> from
    R.P. Monti and P.T. Silvermann "Wastewater system
    Alternates. What are they --- and what cost?"
    Water and Wastes Engineering May 1974.

The costs shown in Table 3 are not prohibitive in light of the dramatic improvement in effluent quality available.

As has been concluded, it is feasible to produce a high quality effluent from this modified lagoon system at a reasonable cost.

Adoption of this system should allow further expansion of municipalities located on marginal receiving streams.

A PRELIMINARY REPORT ON PILOT PLANT FILTRATION OF SECONDARY EFFLUENTS

by

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INTRODUCTION

Recent critical evaluations on assimilative capacities of receiving streams has led to the imposing of high effluent quality objectives for discharges from some Water Pollution Control Plants. This higher quality effluent is frequently described as having the following chemical characteristics: 5 mg/l BOD, 5 mg/l SS, and total phosphorus concentrations of l mg/l or less. Frequently, design considerations include proposals for filtration of secondary effluents (effluent polishing) which are assumed to be capable of yielding the above-mentioned high quality effluent.

Previous pilot studies (1) conducted by the Ministry of the Environment at the Brampton-Chinquacousy WPCP indicated that the effluent quality from a filtration unit is not a function of hydraulic loading or of media type but is predominantly influenced by the filter influent quality. The results of that study indicated that filtration of secondary effluents would not consistently yield the high effluent quality that has been required in some installations.

As effluent polishing is a standard unit operation that can be readily implemented at a WPCP, it was necessary to determine whether an effluent with BOD and suspended solids below 5 mg/l could be consistently obtained through filtration, possibly with the addition of low concentrations of prime coagulants and/or polyelectrolyte to the filter influent.

Since phosphorus removal is now required at most WPCP's in the Province of Ontario, an effort was made to determine the quality of effluent obtainable through filtration of such chemically treated effluents. The characteristics of filter operation under these conditions were assessed simultaneously.

In order to partially achieve these objectives, a pilot filtration unit was installed at the Waterloo WPCP in late 1973 and tests were carried out over 1974. This plant could be considered an average activated sludge plant showing effluent BOD's and SS of 18 and 20 mg/l, respectively. Ferric chloride is added at the end of the aeration tank at a dosage rate of 7 mg/l as Fe^{+3} to effect phosphorus removal. Over 1974 the plant operated at approximately 98% of its hydraulic load and at approximately 83% and 88% of the design load for BOD and suspended solids, respectively.

EXPERIMENTAL PROCEDURE

Feed water for the pilot filter was supplied from the final effluent chamber prior to chlorination via the plant's defoamer and hydrant system. A 300 gallon constant overflow tank for feed water was installed ahead of the pilot unit in order to ensure a continuous supply of final effluent without any intermittent changes in effluent quality.

A flow diagram of the filter unit is presented in Figure 1.

The filter chamber was 22 inches in diameter by 96 inches high with an area of 2.5 sq ft. The media consisted of a top layer of 36 inches of anthracite (0.9 mm effective size, uniformity coefficient 1.65) and a bottom layer of 12 inches of sand (0.5 mm effective size, uniformity coefficient 1.5) supported on gravelless underdrains.

Water was pumped down through the filter and through a "Kates" flow control valve to a backwash storage tank, from which the overflow flowed to waste.

All filter runs were terminated at approximately 60 inches of water headloss above the initial dynamic headloss on the clean filter bed. Hydraulic loads of 2, 5 and 8 Igpm/sq ft were applied to the filter with the majority of the tests carried out at 5 Igpm/sq ft.

Total differential pressure across the filter and the flow were constantly measured and recorded automatically throughout each run.

At the end of Run #8, piezometer taps were installed in the filter column at 6 inch intervals from below the underdrains to above the media and connected via a switching device to a 0 - 200 inches of water differential pressure gauge, thus allowing the measurement of

FIGURE 1. PILOT PLANT FLOW DIAGRAM

LEGEND

⋈	GATE VALVE
	CHECK VALVE
Ⓜ	MOTOR OPERATED VALVE
	RELIEF VALVE
FI	FLOW INDICATOR
FT	FLOW TRANSMITTER
FR	FLOW RECORDER
PI	PRESSURE INDICATOR
DT	DIFFERENTIAL INDICATOR TRANSMITTER
DR	DIFFERENTIAL PRESSURE RECORDER

INLET PUMP

AIR BLOWER

BACKWASH PUMP

BACKWASH TANK
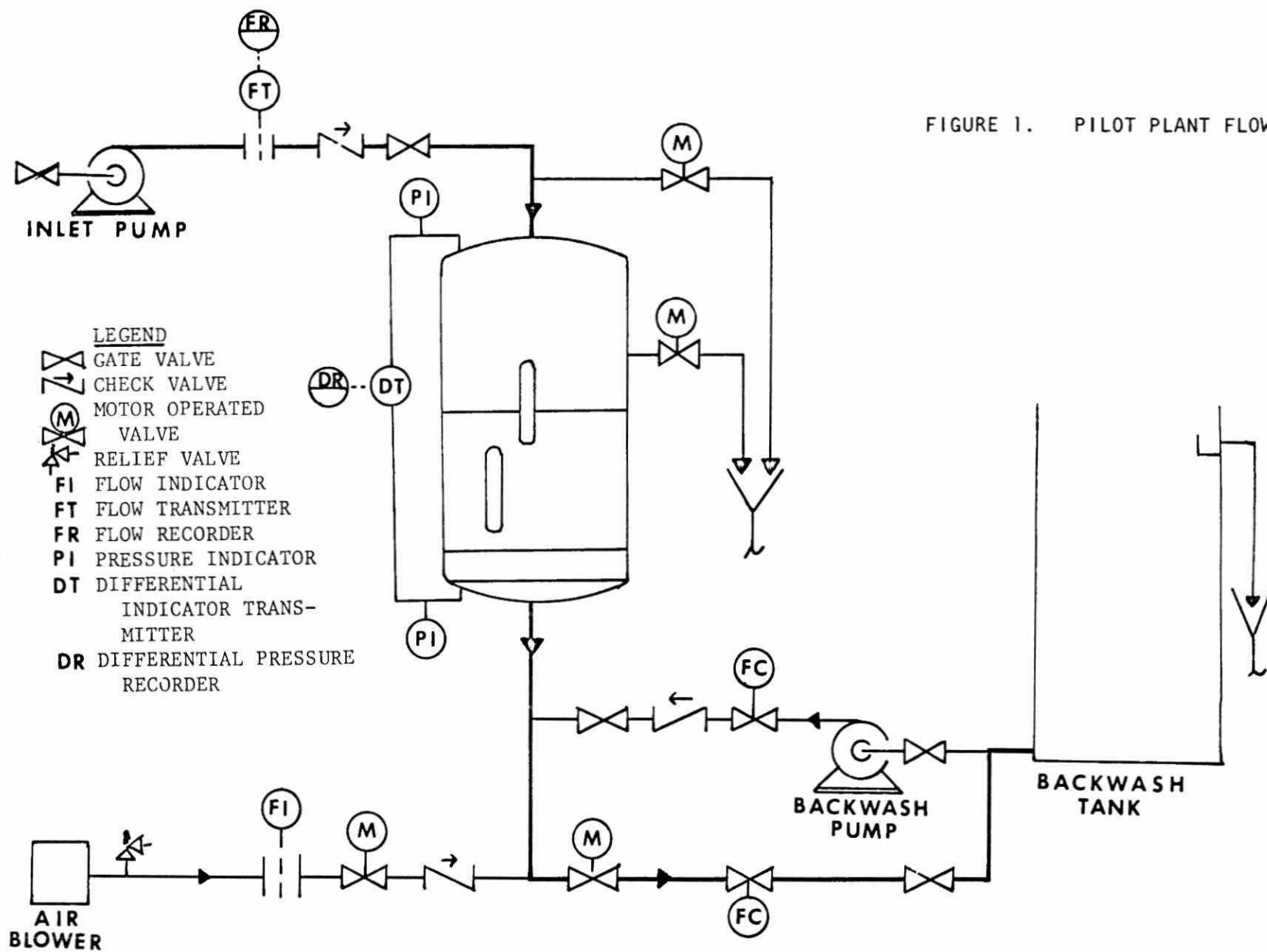
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differential pressure throughout the media. Measurements were taken at the beginning and end of each run as well as at intervals during the run.

Backwash was accomplished automatically when the filter reached 60 inches of headloss above initial dynamic headloss. The backwash sequence consisted of a 5 minute draindown time, 5 minutes of air scour (15 SCM @ 5 psi), and a one minute refill period followed by 5 minutes of backwash at 13 Igpm/sq ft.

Particle size analysis using a Millipore MC Unit was carried out on feed water and filter effluent from one filter run at each flow rate.

Twenty-five runs at various flow rates were carried out in total. Not all runs were taken to completion due to mechanical difficulties either within the filter or within the WPCP.

EXPERIMENTAL RESULTS & DISCUSSION

The experimental results are presented in Appendix 1.

One typical run at each flow rate is presented and discussed in detail. The criteria for choosing the typical runs was that each have a similar and consistent feed water suspended solids range. The typical runs are not necessarily the most efficient runs from each series.

Slow Rate Filtration

Two filter runs were carried out using a flow rate of 2 Igpm/sq ft (Run #18 and #19 in Appendix 1). Run #18 exhibited atypical results in that the feed water suspended solids average of 17 mg/l (range - 12-23 mg/l) was reduced to only 13 mg/l (range - 9-17 mg/l) with the filter only retaining an average of 4 mg/l (range - 3-7 mg/l). Also, bulking in the final clarifier caused the feed water suspended solids to go to 676 mg/l in the nineteenth hour causing the run to quickly go to terminal headloss.

Run #19 began at an initial dynamic clean bed headloss of 20 inches of water and terminated at 81.5 inches of water headloss 28 hours later.

Cumulative total headloss and cumulative solids retention for Run #19 are presented in Figure 2.

The feed water suspended solids average of 12 mg/l (range - 9-15 mg/l) was reduced to an average filter effluent of 3 mg/l (range - 2-5
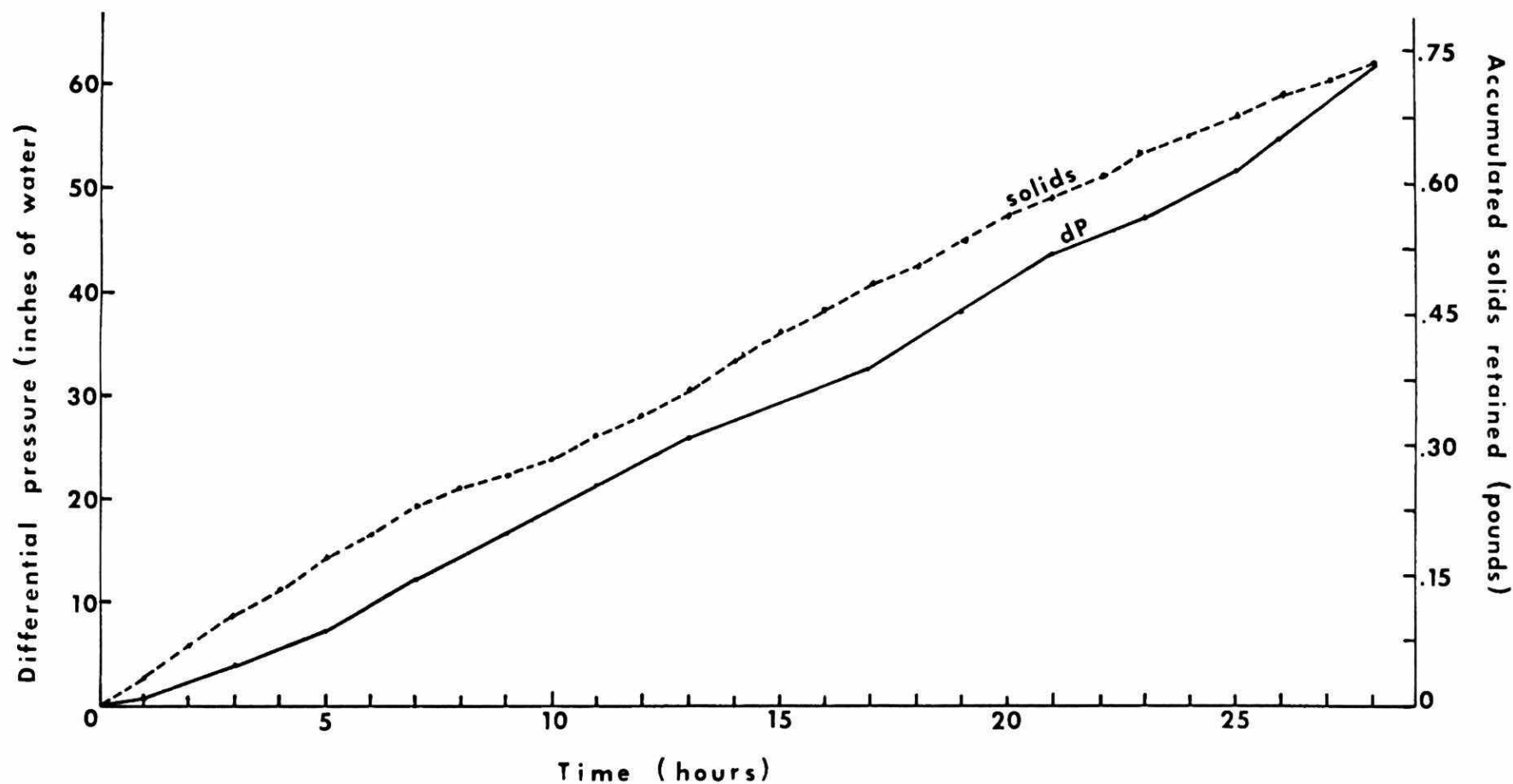
FIGURE 2. DIFFERENTIAL PRESSURE AND SOLIDS RETENTION
FOR PILOT FILTER RUN #19 AT 2 lgpm/sq ft

mg/l) SS with the filter retaining an average of 9 mg/l (range - 6-12 mg/l) suspended solids.

BOD results for Run #19 exhibited similar performance to those for suspended solids, with the average BOD of 18 mg/l (range - 16-21 mg/l) in the feed water being reduced to an average of 4 mg/l (range - 3-4 mg/l) in the filter effluent with the filter retaining an average of 15 mg/l (range - 13-17 mg/l) BOD.

Medium Rate Filtration

Eighteen runs were carried out at 5 lgpm/sq ft. The results from these runs are presented in Appendix 1. Run #10 was chosen as a typical run as it fulfilled the criteria noted above. The run began at an initial dynamic headloss of 45 inches of water and terminated 12 hours later at a headloss of 100 inches of water.

The feed water suspended solids averaged 12 mg/l (range - 9-14 mg/l) and the effluent SS averaged 3 mg/l (range - 1-6 mg/l). BOD values followed a similar pattern, with feed water BOD averaging 16 mg/l (range - 13-20 mg/l) and effluent BOD, 3 mg/l (range - 3-4 mg/l).

Cumulative total headloss and cumulative solids retention for Run #10 are presented in Figure 3.

High Rate Filtration

Five filter runs were carried out at 8 lgpm/sq ft (Appendix 1). Run #22 is presented in detail.

Run #22 began at an initial dynamic headloss of 100 inches of water and terminated six hours later at 157 inches of water.

Influent suspended solids averaged 15 mg/l (range 11-17 mg/l) while the effluent SS were consistent at 4 mg/l SS. BOD levels in the influent feed water averaged 23 mg/l (range - 20-26 mg/l) with an effluent BOD average of 9 mg/l (range - 8-10 mg/l).

Cumulative total headloss and cumulative solids retention for Run #22 are presented in Figure 4.

"Typical" Runs Comparison

The effect of flow rate variation on the quality of effluent produced between the three "typical" runs was not evident in the effluent suspended solids analysis (Appendix 1). A consistent high quality effluent
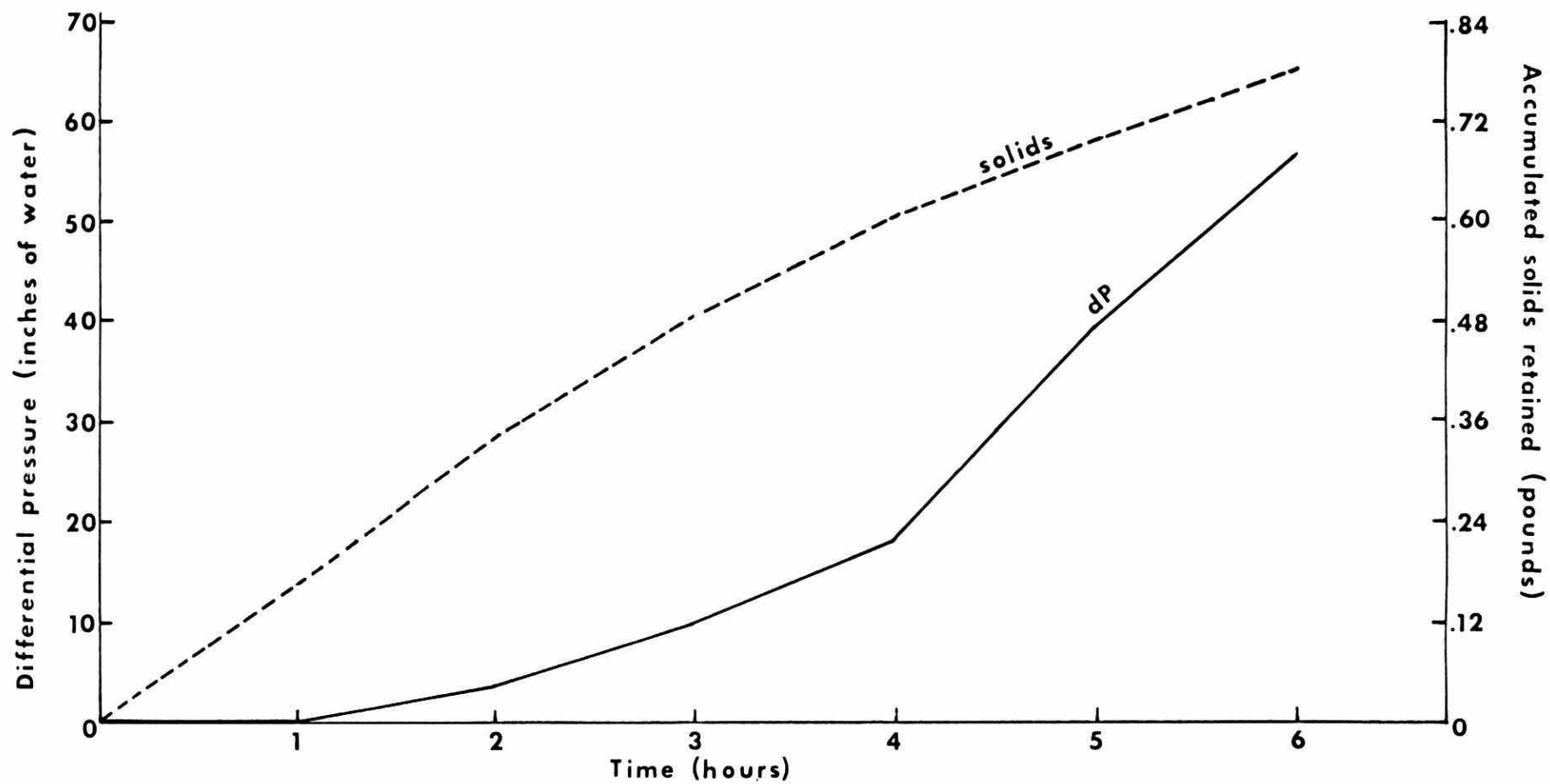
FIGURE 3. DIFFERENTIAL PRESSURE AND SOLIDS RETENTION
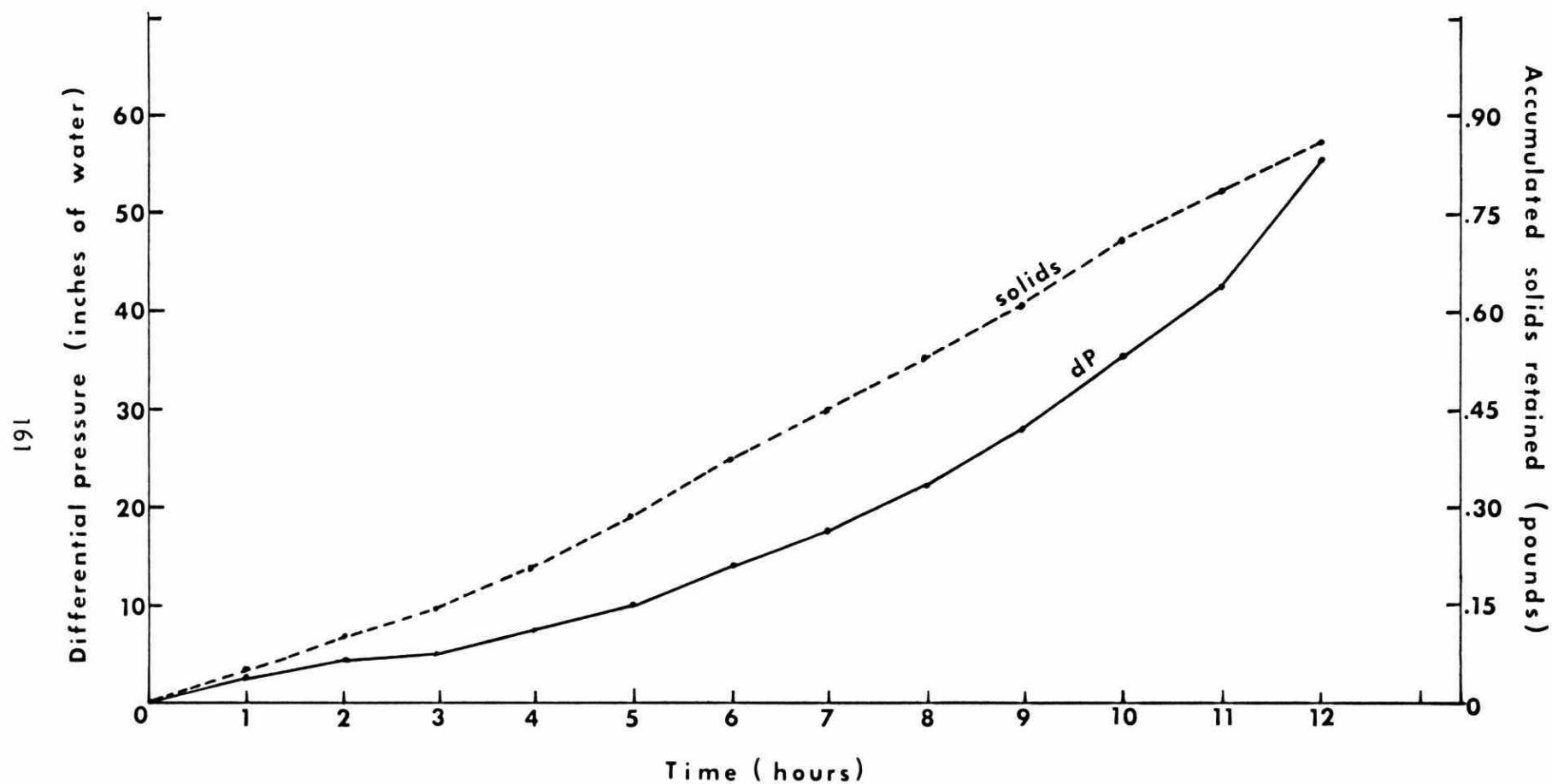FOR PILOT FILTER RUN #10 AT 5 lgpm/sq ft

FIGURE 4.   DIFFERENTIAL PRESSURE AND SOLIDS RETENTION
FOR PILOT FILTRATION RUN #22 AT 8 Igpm/sq ft

of 3-4 mg/l suspended solids was produced. Although the average effluent suspended solids was 1 mg/l higher in the highest flow rate as compared with the two lower flow rates, this difference was not felt to be significant.

The three "typical" runs were examined for variation in effluent suspended solids in relation to variations in influent solids levels (Appendix 1). In all three runs there was no evident relationship, with the effluent SS remaining relatively constant while the influent SS varied. This point is even more evident in runs #8 and #9 (Appendix 1) where intermittent high influent SS levels did not vary the effluent SS levels.

Examination of the effluent SS levels in runs #19 and #10 (2 and 5 lgpm/sq ft) (Appendix 1), shows an improvement of the SS removal ability of the filter over the duration of the filter runs. This would suggest that the trapped solids were reducing the effective size of the media and thus improving suspended solids removal. This phenomena was not evident in the high rate filtration run.

BOD effluent levels in the 2 and 5 lgpm/sq ft "typical" runs were an average of 4 and 3 mg/l, respectively. In the 8 lgpm/sq ft run, the effluent BOD averaged 9 mg/l. The BOD-suspended solids ratio for the influent was 1.5 which is much higher than normally would be expected. Given well stabilized activated sludge plant effluent with a normal BOD-suspended solids ratio in the range of 0.5, the 4 mg/l average suspended solids level in the high rate run would be adequate to reduce the BOD level below 5 mg/l.

The amount of effluent produced per square foot of media for the duration of each of the "typical" runs was calculated. The filter runs at 2, 5 and 8 lgpm/sq ft produced 3360, 3600 and 2880 lgal of effluent per square foot, respectively.

It would be expected that the solids storage in the filter (Figures 1, 2 and 3) would show a similar pattern. The filter runs at 2, 5 and 8 lgpm/sq ft showed 0.74, 0.86 and 0.78 pounds of stored solids/sq ft of media, respectively, at run termination.

The discrepancies in order between the effluent production and the solids storage values are probably a result of the limited data examined. For the available data, it can be concluded that the filter flow rate (lgpm/sq ft) has no influence on the total effluent (gal/run) or solids retention capability (lb SS retained/ft^2 media), to a given terminal headloss above initial dynamic headloss.

Headloss Distribution

For filter runs numbers 9 through 25 a headloss distribution chart is presented in Appendix II. The data presented are those taken immediately prior to terminal headloss. Hutchison (2) states that: "During the course of a filter run, the headloss will distribute itself in a manner similar to that measured at the conclusion of run". Numerous tests during this study confirm this point. The data in Appendix II represent the headloss measured near termination minus the initial dynamic headloss for that same region in the filter taken at the beginning of the run.

The headloss distribution and presumably the solids distribution within the filter at the termination of the three "typical" runs at various flow rates is presented in Table 1. Runs #19 and #10 at 2 and 5 lgpm/sq ft respectively show a similar pattern of distribution, with 72% of the solids stored in the top 3 inches of anthracite below the anthracite-water interface, and 8% and 7% of the solids stored in the next 9 inches of anthracite. The remainder of the anthracite shows little or none of the stored solids. Run #19 exhibits no storage in the sand layer while Run #10 shows 6% in this area but not at the interface. Both runs exhibit high headloss over the underdrain area.

Run #22 (8 lgpm/sq ft) has 77% of the solids trapped in the top 3 inches of anthracite while the area from here to below the sand-anthracite interface exhibits no storage at all. This would suggest that the high flow rate is causing surface blinding at the water-anthracite interface. The sand area between 4 and 10 inches above the underdrain has 9% of the stored solids. Again, a significant headloss is shown across the under-drains.

The lack of solids storage at the anthracite-sand interface
may be a result of an insufficient backwash rate, which would cause a
mixing of the coal and sand and thus negate the effectiveness of the
interface. The solids storage in the sand below the interface in Runs
#10 and #22 would support this theory. This mixing of sand and anthra-
cite may have been advanced to the point where the effective interface
area was moved down to below the 4 inch piezometer point, and the differ-
ential pressure increase noted across the underdrains may have been due
to a lowered interface.

Particle Size Analysis

Three consecutive filter runs were carried out at flow rates
of 2, 5 and 8 Igpm/sq ft and the filter feed water and effluents were
subjected to particle size analysis on a "Millipore πMC Particle Measure-
ment System" at the Canada Centre for Inland Waters (Table 2).

In order to insure the comparability of these samples, suspended
solids analyses were carried out throughout the three runs. Over the
duration, the feed water SS ranged from 10-12 mg/l and the effluent SS
ranged from 3-4 mg/l thus insuring comparability.

The data in Table 2 indicate that the lower the flow rate,
the fewer number of particles found in the effluent. This would be
expected since the higher flow rates would tend to force more particles
through the filter.

The "πMC Particle Measurement System" measures the horizontal
chord of the particles. In order to get another view of these data the
horizontal chord was assumed to be the diameter of a sphere and a "volume"
was calculated. The "volume" of one particle with a diameter of 91.3 microns
is equivalent to the "volume" of approximately 16,000 particles with a
diameter of 3.6 microns. Figure 5 presents the cumulative percent of
total "volume" retained by the filter over the range of particle sizes
for the three flow rates used. Also, cumulative percent of total "volume"
of the raw water is presented.

The particle count (Table 2) shows 85% of the number of particles
in the feed water below 18.1 microns in chord width. The graph (Figure 5)
suggests that in terms of "volume" this represents approximately 12%

TABLE 1.   HEADLOSS DISTRIBUTION AT VARIOUS
FLOW RATES

Filter Depth In	Run #19 2 lgpm/sq ft %	Run #10 5 lgpm/sq ft %	Run #22 8 lgpm/sq ft %
-49			
-43	72	72	77
-40	6	3	(-2)
-34	2	4	2
-28	1	0	0
-22	1	(-1)	0
-16	1	(-1)	0
-10	2	0	(-4)
- 4	(-2)	6	9
(-2)	19	17	18

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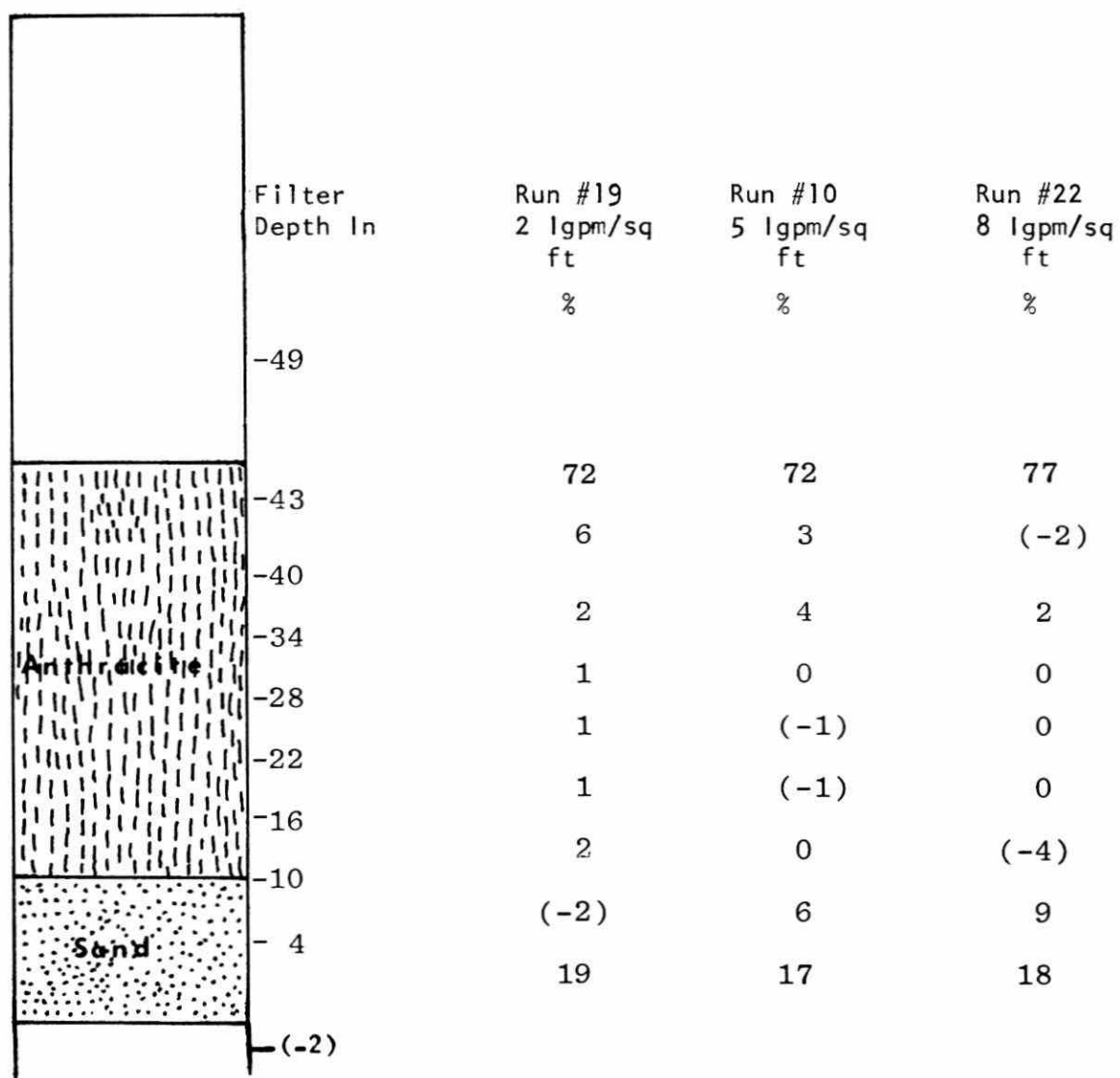
TABLE 2. MILLIPORE πMC PARTICLE SIZE ANALYSIS FOR
PILOT PLANT STUDY

Mean Number of Particles *

Size Mean Horizontal Chord in Microns	Feed Water	Effluent at Various Flow Rates		
		2 lgpm/ sq ft	5 lgpm/ sq ft	8 lgpm/ sq ft
3.6	163	184	111	265
4.5	335	163	253	430
5.7	185	107	118	174
7.2	118	115	159	104
9.1	165	54	106	56
11.4	80	32	98	49
14.4	144	23	71	31
18.1	66	3	14	25
22.8	59	6	40	24
28.7	65	4	11	11
36.2	24	4	7	6
45.6	13	2	3	3
57.4	10	0	1	0
72.3	3	0	0	0
91.3	1	0	0	0
115	0	0	0	0
Total Number	1,431	697	992	1,178

* all data given are an
average of 3 samples

FIGURE 5.   ACCUMULATED "VOLUME" OF PARTICLES EXAMINED IN
THE PILOT PLANT FILTRATION STUDY
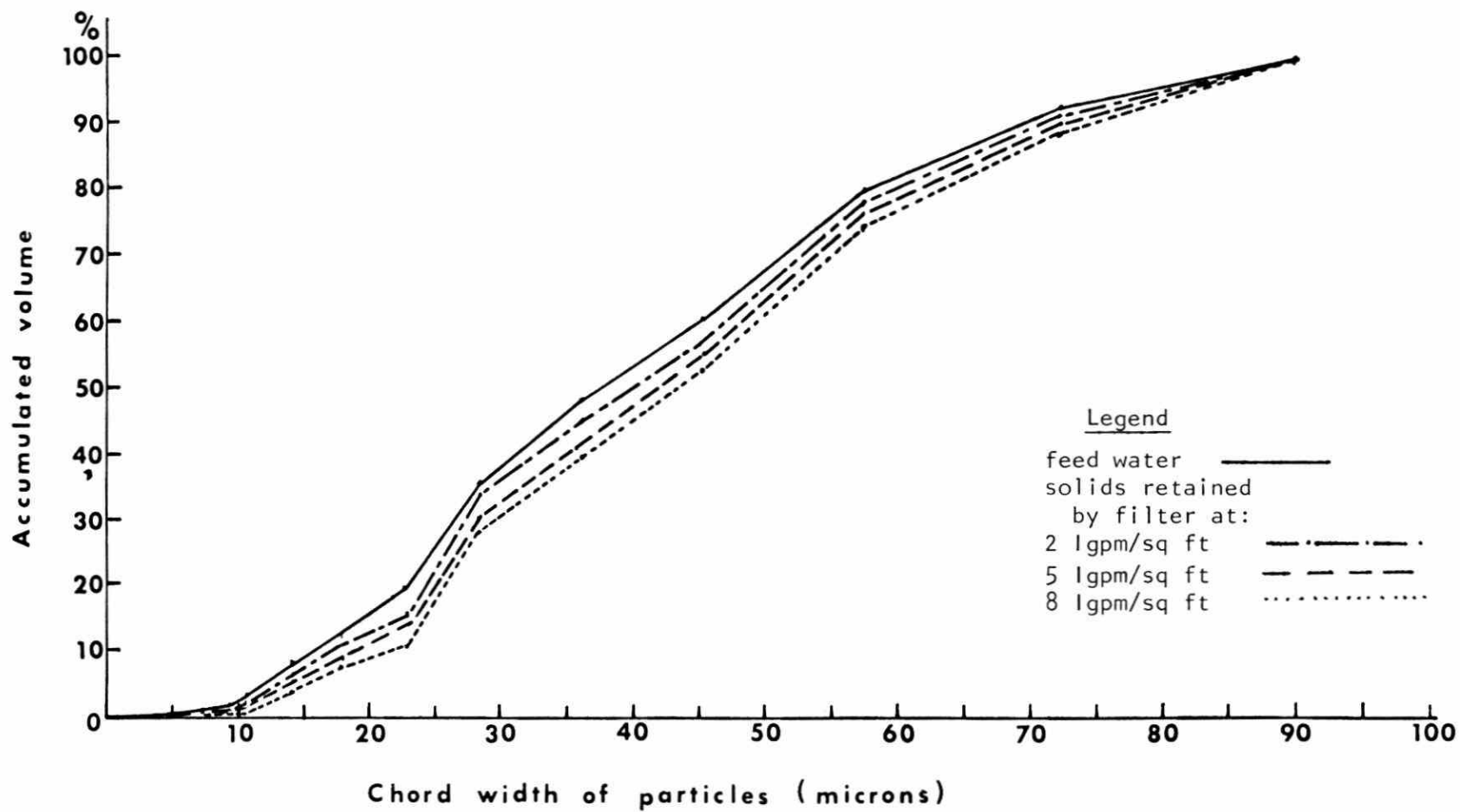
of the "volume" of the feed water. Similarly, the particles below 18.1 microns represent approximately 10% of the total volume retained by the filter.

At the high flow rate (8 lgpm/sq ft) it appeared that particle shear was taking place, since there was an increase in the number of particles in the 3.6 and 4.5 micron size in the effluent compared to the influent samples. Although this only represented a very small fraction of the suspended solids in the effluent, it is anticipated that this adverse effect may become more significant if the hydraulic rate is increased or if the influent suspended solids are less stable.

The efficiency of the filter in terms of particle removal is much greater for particles with a chord width greater than 14.4 microns (Table 2). In this study these particles represented a large percentage of the volume. In the treatment of an effluent where a high percentage of the particles are below 14.4 microns chord width, coagulation would be necessary.

Backwash Values

At flow rates of 2, 5 and 8 lgpm/sq ft the pilot filter unit produced 300, 750 and 1,200 lgal/hr, respectively. A comparison of the gallonage of product over each filter run compared with the backwash volume of 200 lgal will give an index of filter efficiency.

The average percentage of product water used for backwash was calculated for the three flow rates.

Flow Rate lgpm/sq ft	Number of Runs	% of Product Used for Backwash	
		Average	Range
2	2	3.0%	2.4 - 3.5%
5	9	3.1%	1.6 - 4.5%
8	3	3.4%	2.8 - 4.2%

With a variation of 3.0 to 3.4% of product used for backwash, the pilot filtration unit would seem to be quite efficient hydraulically, considering that in some of these runs the feed water had intermittent high SS which ended the filter run much sooner than usual.

Filtration Costs Estimate

A cost estimate based on the ENR Index 2200 (3), January 1973 is presented below. The capital amortization period is 20 years at an interest rate of 8%.

	1.0 MGD WPCP	3.0 MGD WPCP
Capital Costs (4)	190,000.	365,000.
Capital Cost Amortization (20 yr @ 8%) ($/MG)	51.75	33.14
Operation & Maintenance Costs ($/MG)	100.80	72.00
Total Costs ($/MG)	152.55	105.14

SUMMARY AND CONCLUSIONS

1. From the comparison of three runs at 2, 5 and 8 lgpm/sq ft flow rates, the following conclusions were drawn:

   a) Effluent suspended solids did not vary directly with influent suspended solids. A consistently high quality effluent was produced with suspended solids below 5 mg/l.

   b) The three flow rates produced similar effluent quality with average suspended solids levels of 3-4 mg/l.

   c) The two lower flow rates exhibited improving solids removal ability over the duration of the runs. This was not evident in the high rate filtration run.

   d) The two lower flow rates had effluent BOD levels below 5 mg/l. The higher flow rate had an average BOD of 9 mg/l. A high BOD-suspended solid ratio suggested that modification of the secondary treatment process would reduce the 9 mg/l to within acceptable limits.

   e) No significance was given to the variation in the amount of effluent produced by the filter or the solids storage capacity between the three "typical" runs.

f) The ability of the filter to take high influent suspended solids for a short period and deliver consistently low effluent suspended solids would offer a good form of secondary protection for short term plant upsets.

2. a) Particle size analysis showed that the filter was more efficient in the removal of particles greater than 14.4 microns in chord width. Thus, in situations where a high percentage by volume of the particles were below 14.4 microns, coagulation of the filter feed would be required.

   b) Particle size analysis appears to show a shearing of solids into smaller particles at the high flow rate (8 Igpm/sq ft).

3. The majority of the removal of solids in the filter took place at the anthracite-water interface.

4. Hydraulic efficiency of the pilot unit was good with approximately 3.2% of the effluent used for backwash.

5. The costs are not prohibitively high, with the 1 and 3 MGD plants showing total costs of $152.55/MG and $105.14/MG, respectively.

REFERENCES

(1) Rupke, J.W.G., "Sand Filtration". Presented at Pollution Control Association of Ontario Annual Meeting, Ottawa, Ontario, April 14, 1972.

(2) Hutchison, W.R., Operation Variables and Limitations of Direct Filtration, Ontario Ministry of the Environment, Toronto, Ontario, January, 1975.

(3) Monti, R.P. & Silbermann, P.T., "Waste Water System Alternates: What are they and what cost?" Water & Waste Engineering, p 40, May, 1974.

EXPERIMENTAL RESULTS OF PILOT FILTRATION
STUDIES AT WATERLOO 1974

Run Number	Flow Rate Igpm/ sq ft	Time in hours	Differential Pressure (inches of water)	Suspended Solids		BOD	
				Raw (mg/l)	Eff. (mg/l)	Raw (mg/l)	Eff. (mg/l)
1	5	0	58.5				
		1	61.5	10	3		
		2	69.0	7	3		
		3	82.0	6	3		
		4	88.0	6	3		
		5	92.5	6	3		
2	5	0	46	8	3	7	4
		1	49	8	4		
		2	50	8	4		
		3	52.5	8	4	8	4
		4	61.0	10	4		
		5	75.0	10	4	11	4
		6	100.0	11	4		
3	5	0	46.5				
		1	48.5	17	4	9	3
		2	52.5	11	3		
		3	62.0	11	3	9	3
		4	75.0	14	4		
		5	95.0	15	3	10	4
4	5	0	45				
		1	48.5	17	4	14	7
		2	52.5	19	4		
		3	63.0	27	4	25	9
		4	100.0	227	4	100	9
5	5	0	47.5				
		1		226	8	124	11

Run Number	Flow Rate	Time in hrs.	Differential Pressure	Suspended Solids		BOD	
	Igpm/ sq ft		(inches of water)	Raw (mg/l)	Eff. (mg/l)	Raw (mg/l)	Eff. (mg/l)
6	5	0		316	9	151	11
		1					
7	5	DELETED					
8	5	0	45.0				
		1	50.0	16	4	14	2
		2	55.0	15	4		
		3	61.0	14	3		
		4	66.0		5		
		5	71.0	35	5		
		6	100.0	11	6	11	3
9	5	0	47.0				
		1	47.5	16	3	17	4
		3	49.	12		19	3
		4	53.0	42	3		
		5	55.0	11	3		
		6	67.0	23	3	32	5
		7	91.0	67	3		
10	5	0	45.0				
		1	47.5	12	6	13	3
		2	49.0	12	4		
		3	50.0	9	4		
		4	52.5	11	2	15	3
		5	55.0	12	2		
		6	59.0				
		7	62.0	12	1		
		8	67.0	12	2		
		9	72.5				
		10	80.0	14	2	18	3
		11	87.0	14	2		
		12	100.0	11	2	20	4
11	5	0	42.5				
		1	52.0	14	5	18	5
		2	59.0	10	4		
		3	63.0	11	4		
		4	74.5	16	5	18	4
		5	82.5	15	4		

Run Number	Flow Rate lgpm/ sq ft	Time in hrs.	Differential Pressure (inches of water)	Suspended Solids		BOD	
				Raw (mg/l)	Eff. (mg/l)	Raw (mg/l)	Eff. (mg/l)
12	5	0	47.5				
		1	49.0	10	3	14	5
		2	52.0	11	2		
		3	57.5	12	2		
		4	63.5	14	3	19	4
		5	93.0	16	2	22	4
13	5	0	47.0	7	5	18	8
		1	50.0	7	5	18	8
		2	51.0	7	4		
		3	52.5	7	3		
		4	55.0	7	3		
		5	57.5	7	3		
		6	60.0	6	3		
		7	62.5				
14	5	0	48.0				
		1	49.0	9	5	10	6
		2	51.0	7	4		
		3	55.0	7	4		
		4	60.0	8	3		
		5	65.0	9	4	15	5
		6	70.0	10	5		
		7	75.0	9	5		
		8	80.5	10	5		
		9	87.0	11	4	18	5
		10	92.5	10	4		
15	5	0	42.5				
		1	47.5	6	3	7	3
		2	49.0				
		3	50.0	7	3		
		4	50.5				
		5	52.0	7	3		
		6	54.5				
		7	55.0	8	4		
		8	57.5				
		9	60.0	8	3	10	3
		10 ↓					
		14	84.0	13	4		
		15	87.0				
		16	92.5				
		17	100.0	14	5	14	5

| Run Number	Flow Rate	Time in hrs.	Differential Pressure	Suspended Solids		BOD	
	lgpm/ sq ft		(inches of water)	Raw (mg/l)	Eff. (mg/l)	Raw (mg/l)	Eff. (mg/l)
16	5	0	49.0				
		1	49.5	6	2	9	3
		3	54.0	6	2		
		4	56.0				
		5	58.0	4	3		
		6	60.5				
		7	63.0	6	2		
		8	67.0				
		9	70.5	6	3	13	8
		10	74.0				
		11	80.0	8	3		
		12	84.5				
		13	90.0	9	4	18	6
		14	97.5				
		15	100.0				
17	5	0	51.5				
		1	52.0	9	5	16	9
		2	53.5				
		3	57.0	8	5		
		4	58.5				
		5	61.5	7	5		
		6	64.5				
		7	67.0	7	5	13	5
		8	69.5				
		9	74.5	24	4		
		10	86.0				
		11	92.0	7	4	11	5
		12	100.0				
18	2	0	22.0				
		1	22.0	17	10	20	9
		3	22.5				
		5	25.0	19	17		
		7	28.0				
		9	32.0	23	17		
		11	36.5				
		13	42.0	14	11		
		15	48.5	12	9	13	6
		17	55.0				
		19	81.0	676	7	382	8

Run Number	Flow Rate (lgpm/sq ft)	Time in hrs.	Differential Pressure (inches of water)	Suspended Solids Raw (mg/l)	Eff. (mg/l)	BOD Raw (mg/l)	Eff. (mg/l)
19	2	0	20.0				
		1	20.5	15	5	21	4
		3	24.0				
		5	27.5	16	4		
		7	32.5				
		9	37.0	9	4		
		11	41.5				
		13	46.0	13	3	18	4
		15					
		17	52.5	11	2		
		19	58.5				
		21	63.5	11	3		
		23	67.5				
		25	72.0	9	2		
		27					
		28	81.5	9	3	16	3
20	8	DELETED					
21	8	0	97.Q				
		1	98.0	7	4	21	12
		2	100.0				
		3	106.0	9	4		
		4	114.0				
		5		12	2	40	27
22	8	0	100.0				
		1	100.0	18	4	20	10
		2	104.0				
		3	110.0	17	4		
		4	118.0				
		5	140.0	11	4	26	8
		6	157.0				
23	8	0	100.0				
		1	156.0				
24	8	0	106.0				
		1	116.0	43	9	36	19
		2	128.0	32	24		
		3	141.0	52	30		
		4	156.0	156	34	152	32

Run Number	Flow Rate (gpm/ sq ft)	Time in hrs.	Differential Pressure (inches of water)	Suspended Solids Raw (mg/1)	Eff. (mg/1)	BOD Raw (mg/1)	Eff. (mg/1)
25	8	0	109.0				
		1	121.0	7	5	14	6
		2	126.0				
		3	134.0	10	5		
		4	141.0				
		5	146.0	10	5	21	6
		6	156.0				

APPENDIX II

HEADLOSS DISTRIBUTION FOR PILOT PLANT
FILTRATION STUDY

Depth of media (inches)

177
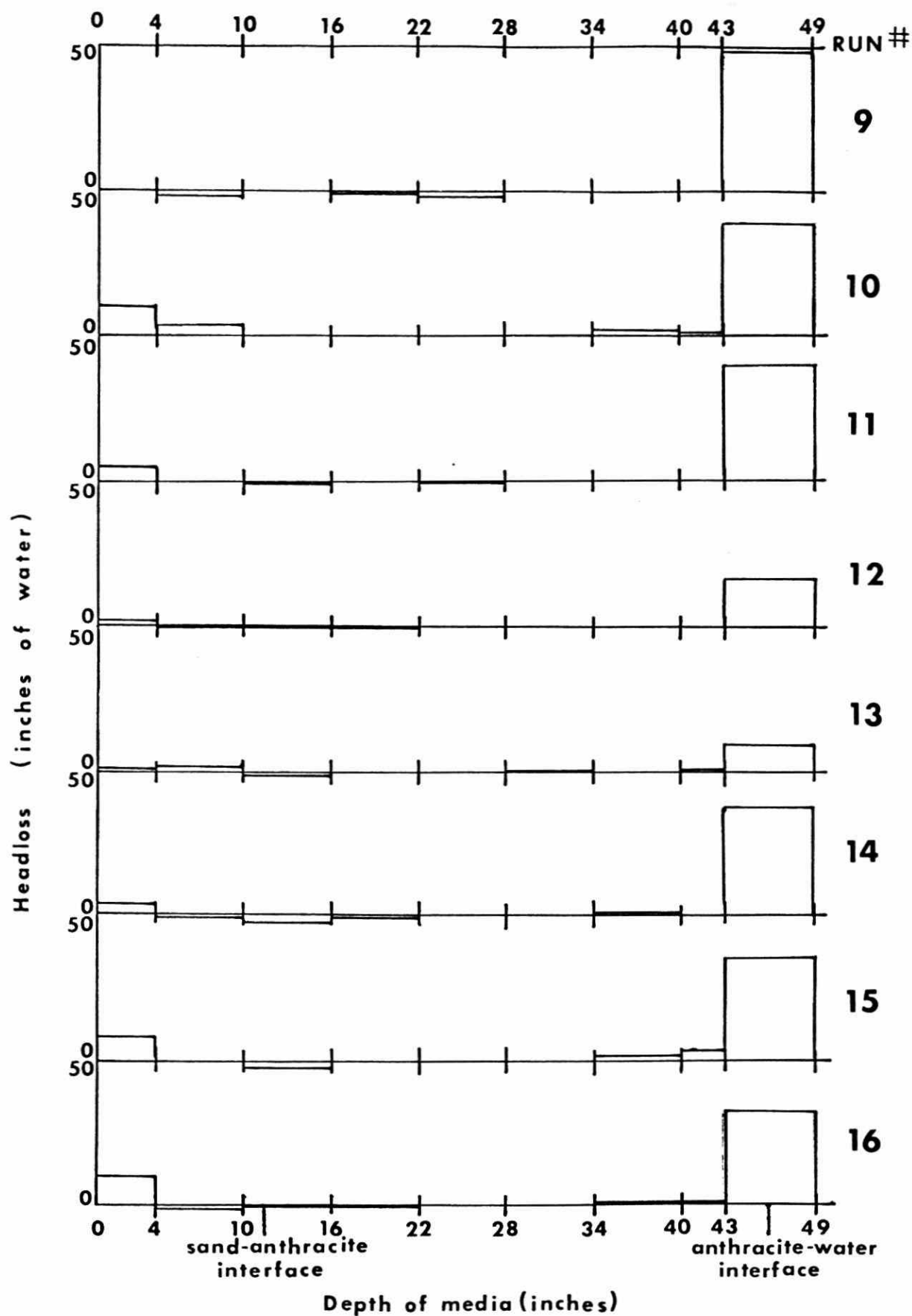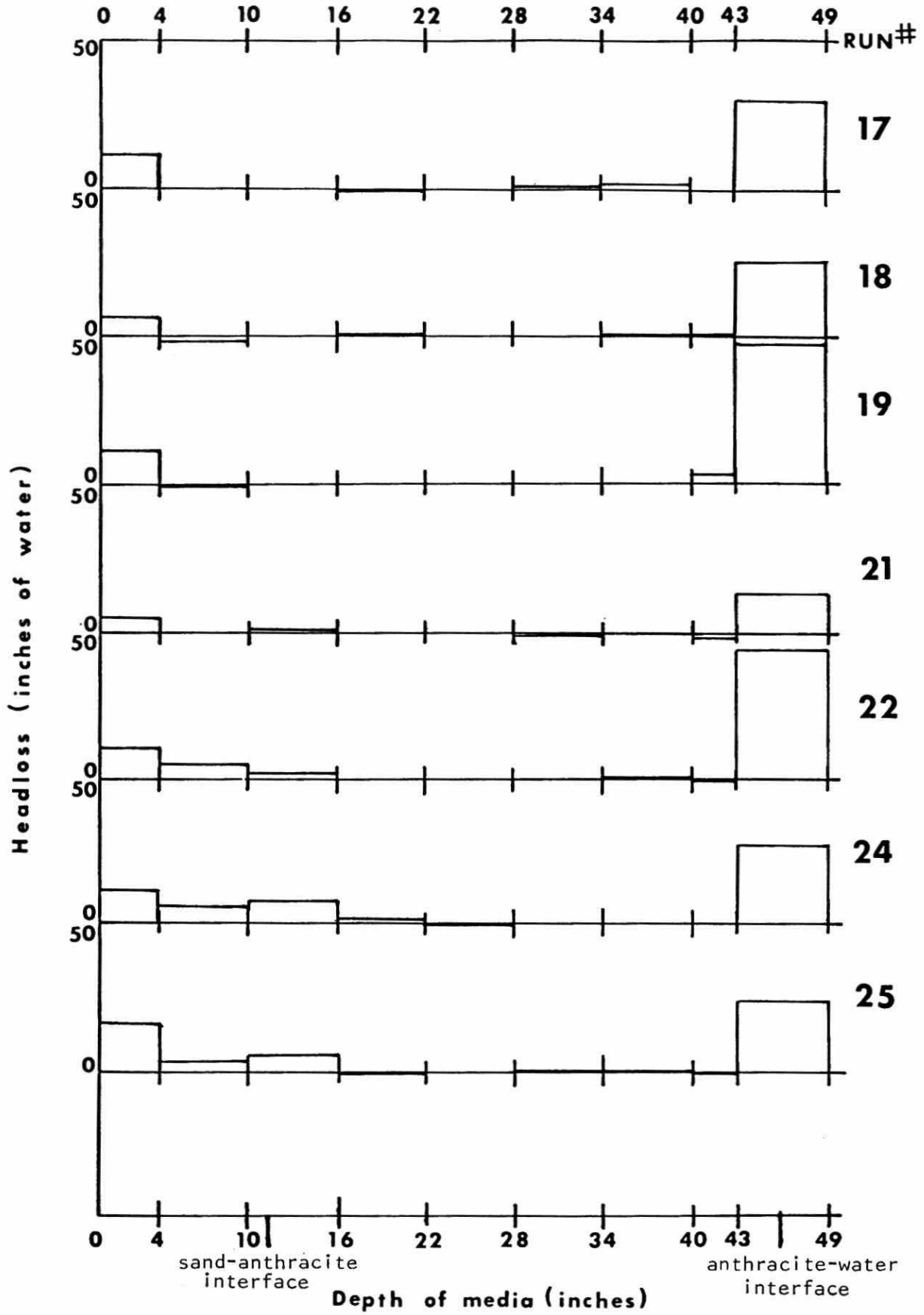
DESIGN CONSIDERATIONS TO ATTAIN LESS THAN
0.3 mg/l EFFLUENT PHOSPHORUS

by

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INTRODUCTION

In response to the Canada/Ontario and the Canada/United States Agreements on Great Lakes Water Quality, the Province of Ontario implemented a program for the installation of phosphorus removal facilities at municipal and institutional wastewater treatment plants located in the Upper and Lower Great Lakes Drainage Basins, the Ottawa River System and inland recreational areas. Today, effluent total phosphorus concentrations not exceeding one milligram per litre are a requirement for the majority of wastewater treatment facilities in Ontario. However, more stringent effluent phosphorus criteria may be required in certain areas in the future because of a sensitivity of the receiving water to phosphorus discharges (Lee, 1970). This paper will discuss design considerations for attaining total phosphorus concentrations less than 0.3 mg/l in the final effluent discharged from water pollution control plants.

Background

To implement the Ontario Government's phosphorus removal policy, the Ministry of the Environment initiated a two-phase treatability study programme consisting of jar testing and short-term full scale studies financed under the Canada/Ontario Agreement (COA). Since the initiation of the program in 1972, phosphorus removal treatability studies have been carried out by personnel of the Ontario Ministry and the Federal Department of the Environment, as well as municipal staff and consulting engineers, at over 200 water pollution control plants in Ontario. A summary of the findings of these treatability studies is contained in an as yet unpublished report by Prested, et al. (1975). The report concludes that effluent total phosphorus concentrations of 1 mg/l can be achieved by traditional coagulation and precipitation processes using lime, or iron or aluminum salts.

A summary of some of the results obtained during these full scale treatability studies in Ontario is presented in Table 1. The optimum method for phosphorus removal in most instances was addition of ferric chloride or alum to the mixed liquor. With mixed liquor chemical addition, the studies showed that, at a final effluent total phosphorus concentration of 1 mg/l, the average metal cation to filterable phosphorus molar ratios were 2.3 and 2.1 for iron and aluminum, respectively. Of the 12 plants that used alum or ferric chloride addition to the raw wastewater, seven were primary plants and had no other option.

FUNDAMENTAL PHYSICO-CHEMICAL CONSIDERATIONS FOR PHOSPHORUS REMOVAL

Phosphorus is generally present in wastewater in more than one chemical form. The relative quantities of the three major forms of phosphorus (orthophosphate, polyphosphate, and particulate phosphorus) generally encountered in domestic wastewater are shown in Figure 1. Approximately one-third of the total phosphorus is in the soluble orthophosphate form, another third in the soluble polyphosphate form, and the remaining one-third is in the form of suspended particulate phosphorus. The final effluent total phosphorus concentration attainable depends upon the degree of conversion of soluble phosphorus to an insoluble form through reaction with a chemical coagulant/precipitant, and the efficiency of liquid-solids separation and suspended solids removal in the treatment plant. To achieve total phosphorus concentrations of 0.3 mg/l or less, virtually complete conversion of soluble phosphorus to particulate phosphorus is necessary. For all practical purposes, the residual total phosphorus concentration in the final effluent then becomes a direct function of the efficiency of the treatment system for suspended solids removal.

It would be appropriate at this time to briefly consider the mechanisms responsible for phosphorus removal by chemical precipitation. The chemical reactions operative in phosphorus removal through the addition of lime are substantially different from those that occur when aluminum or iron salts are employed as precipitants. These coagulants are therefore discussed separately in the following sections.

TABLE 1   SUMMARY OF SOME RESULTS FROM FULL SCALE TREATABILITY STUDIES*

Chemical	Total No. of Plants	Addition to Raw Wastewater				Addition to Mixed Liquor			
		Number of Plants	Avg. Dosage (mg/1)	Avg. Molar Ratio Metal Ion to		Number of Plants	Avg. Dosage (mg/1)	Avg. Molar Ratio Metal Ion to	
				TP[4]	FP[5]			TP[4]	FP[5]
Ferric Chloride	27	7	14.2[1]	2.7	4.2	20	9.5[1]	1.5	2.3
Alum	20	5	10.3[1]	1.7	-[3]	15	7.5[1]	1.6	2.1
Lime	4	4	101[2]	15.8	-[3]	-	-	-	-

*   Effluent total phosphorus objective of 1 mg/1

[1]   Calculated as Fe or Al

[2]   Calculated as Ca

[3]   Insufficient data

[4]   TP = Total Phosphorus

[5]   FP = Filterable Phosphorus
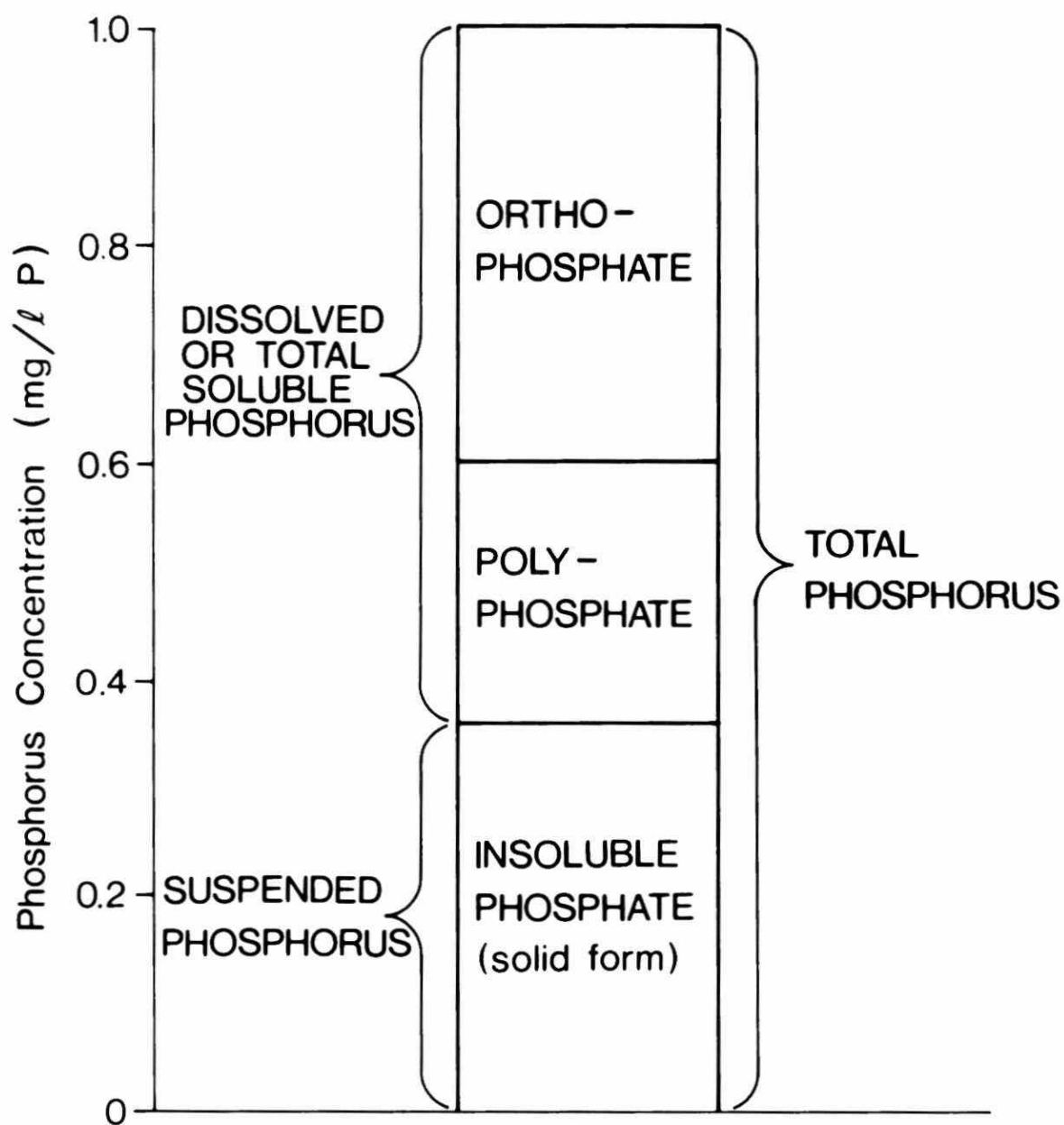
FIG.1 RELATIVE DISTRIBUTION OF PHOSPHATES
IN SEWAGE (SCHUESSLER, 1970)

Aluminum and Iron Precipitation

The mechanisms responsible for the removal of phosphorus from wastewater by the addition of aluminum or iron salts are of a complex nature. They involve a number of competing chemical equilibria dependent upon pH, temperature, alkalinity, hardness and other wastewater characteristics. For more specific information the reader is referred to the literature (Jenkins, et al., 1971; Marson, 1971).

Perhaps the most comprehensive laboratory jar test study using aluminum and iron salts for phosphorus removal has been carried out by Recht and Ghassemi (1970). According to this study, the optimum pH for orthophosphate precipitation from pure solutions was 6.0 for aluminum and about 4.0 for iron, at metal cation to orthophosphate molar ratios in the range of 0.5 to 2.0. Residual orthophosphate concentrations obtained at various pH's are shown in Figure 2 for a 2:1 molar ratio of aluminum or iron to orthophosphate. Orthophosphate concentrations less than 0.3 mg/l, calculated as P, were obtained at pH values of 5.5 to 7.0 for aluminum and 3.5 to 5.5 for iron. Similar results were obtained with aluminum addition to a secondary effluent (Figure 3).

The efficiency of orthophosphate removal was shown to be a function of both operating pH and aluminum/iron dosage. At the optimum pH for orthophosphate precipitation, as well as at slightly lower pH levels, the chemical reactions resulted in the formation of extremely fine and well stabilized solids (as revealed by the appearance of turbidity). In the case of iron, at pH values in the range from 3 to 4, the resultant turbidity could only be removed effectively by filtration through a 0.1 μm membrane. However, within the pH range of approximately 5 to 7 for Al^{3+} and 4 to 6 for Fe^{3+}, large gelatinous precipitates formed and these settled very readily.

Recht and Ghassemi also observed that variations in aluminum and iron dosages had an effect on orthophosphate removal efficiency similar to that just described for pH. Thus, at the optimum pH values complete orthophosphate precipitation was achieved at Fe^{3+} to PO_4^{3-} molar ratios of 1.4 to 1.6, and Al^{3+} to PO_4^{3-} molar ratios of 1.8 to 2.0. Increases in Fe^{3+} to PO_4^{3-} molar ratios above 1.6, and Al^{3+} to PO_4^{3-} to molar ratios above 2.0 resulted in the formation of fine precipitates which did not settle
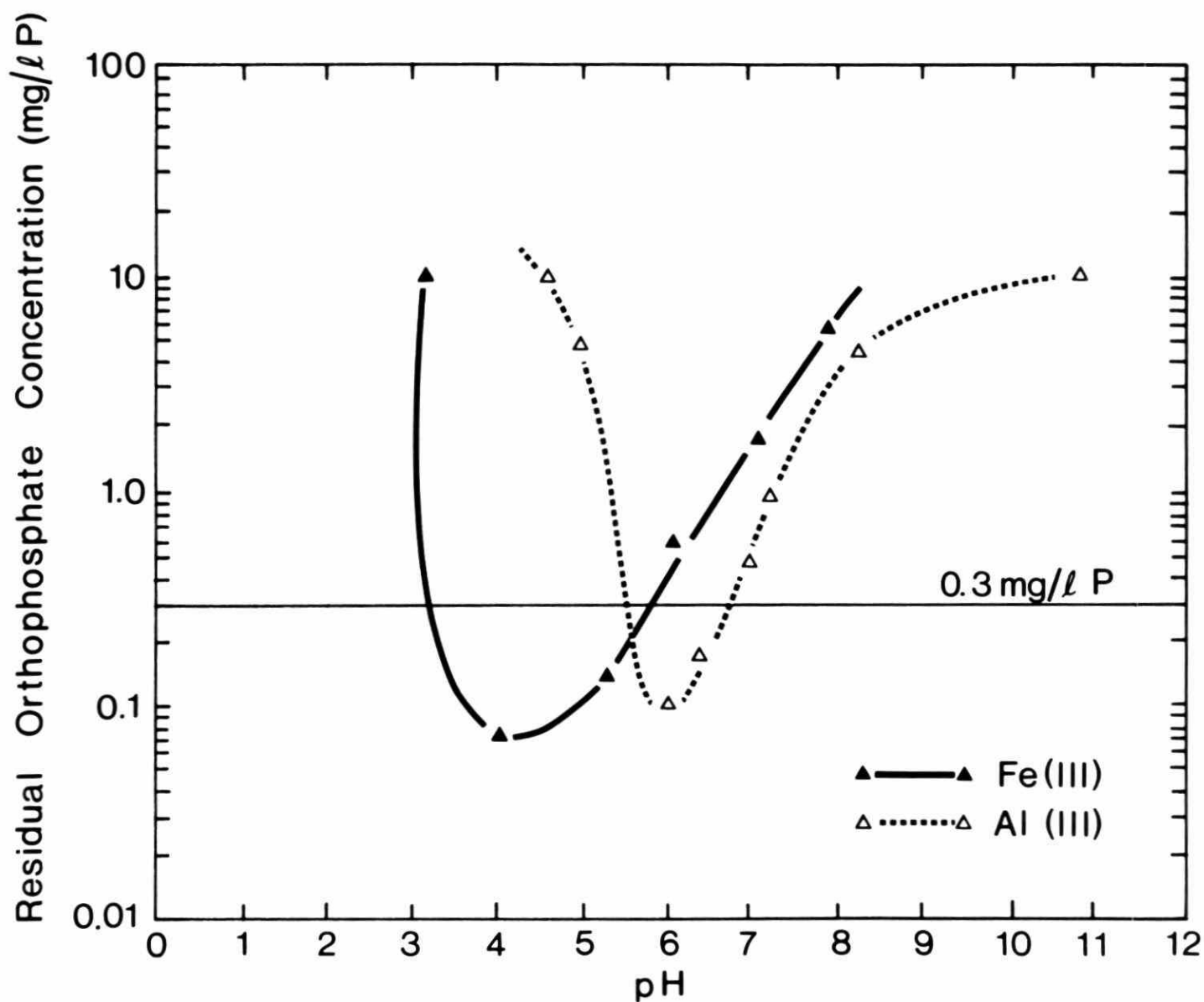
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FIG.2 PRECIPITATION OF ORTHOPHOSPHATE WITH Al (III) AND Fe (III) AT A 2 : 1 CATION - TO - ORTHOPHOSPHATE MOLAR RATIO. (INITIAL ORTHOPHOSPHATE CONCENTRATION, 12 mg/ℓ P) (RECHT AND GHASSEMI, 1970)
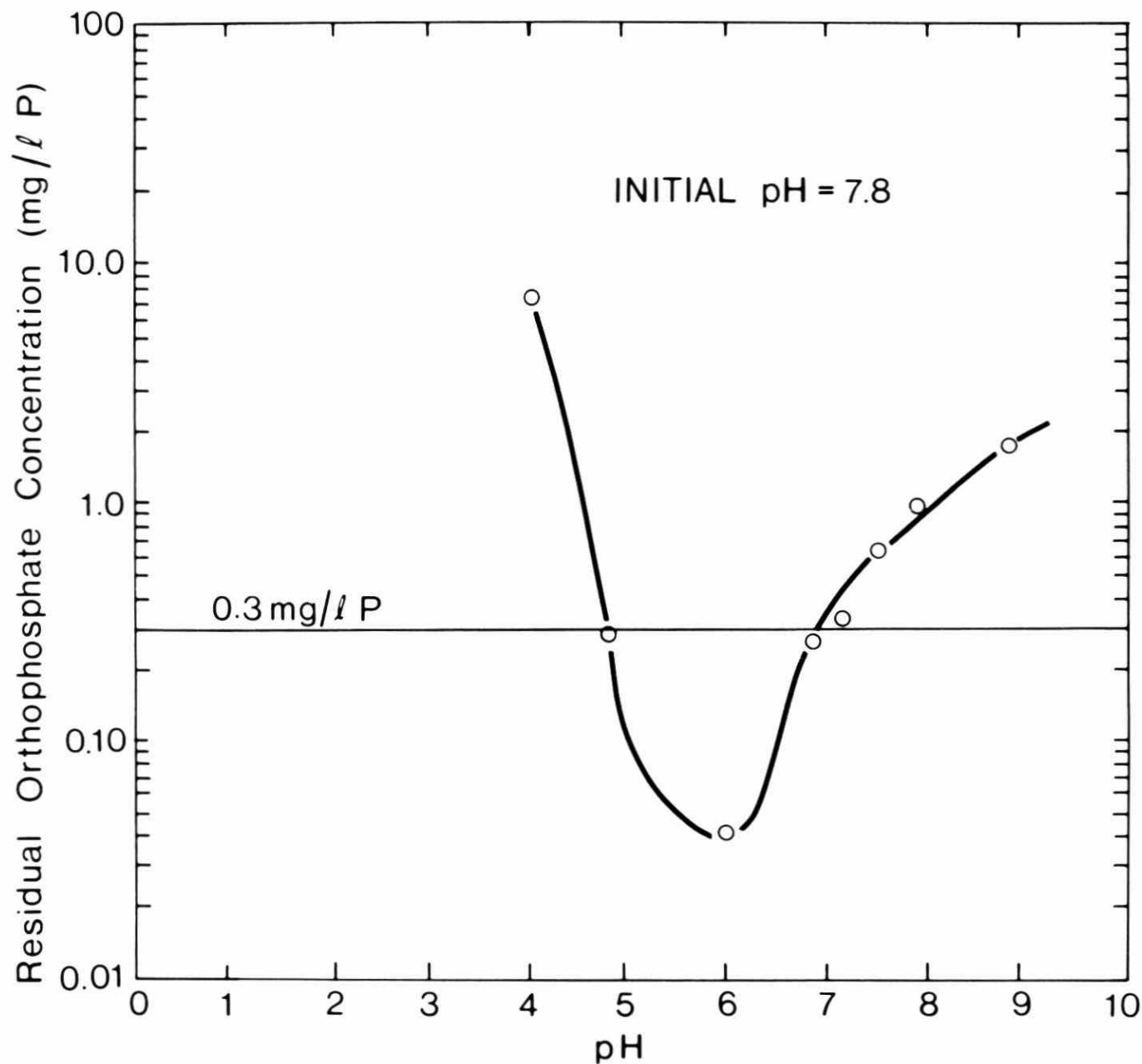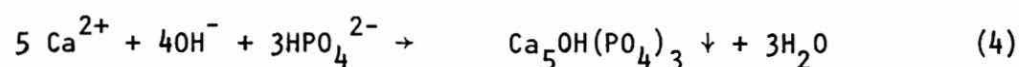
FIG.3   PRECIPITATION OF ORTHOPHOSPHATE FROM SECONDARY EFFLUENT WITH Al(III) AT A 2:1 CATION-TO-ORTHOPHOSPHATE MOLAR RATIO. (INITIAL ORTHOPHOSPHATE CONCENTRATION, 7.75 mg/ℓ P) (RECHT AND GHASSEMI, 1970)

very well. These suspended solids could only be removed through a membrane filter (0.45 μm). Without filtration the increased dispersion of these precipitates resulted in corresponding increases in the metal (Fe or Al) and phosphorus concentrations in the supernatant.

Lime Addition

Phosphorus removal by means of lime addition is strongly pH dependent and thus directly related to the wastewater alkalinity. Chemical reactions of interest during lime addition to wastewater include (O'Farrell and Biship, 1971; Narkis et al., 1975):

$$Ca^{2+} + 2OH^- + CO_2 \rightarrow CaCO_3 \downarrow + H_2O \qquad (1)$$

$$Ca^{2+} + 2OH^- + HCO_3^- \rightarrow CaCO_3 \downarrow + H_2O + OH^- \qquad (2)$$

$$Ca^{2+} + 2OH^- + CO_3^{2-} \rightarrow CaCO_3 \downarrow + 2OH^- \qquad (3)$$

$$5\, Ca^{2+} + 4OH^- + 3HPO_4^{2-} \rightarrow Ca_5OH(PO_4)_3 \downarrow + 3H_2O \qquad (4)$$

$$Mg^{2+} + 2OH^- \rightarrow Mg(OH)_2 \downarrow \qquad (5)$$

Precipitation of magnesium hydroxide (Equation 5) begins above pH 10.5.

The lime dosage required to achieve a specific pH level in a wastewater increases with increasing wastewater alkalinity as shown in Figure 4. As has been pointed out in the previous section, total phosphorus removal through lime addition is a strong function of operating pH. The relationship between residual phosphorus concentration and operating pH for a raw wastewater of moderate alkalinity (100 to 150 mg/l $CaCO_3$) is shown in Figure 5. This illustration shows that, while the total soluble phosphorus concentration is reduced to less than 0.3 mg/l P at pH 10, the total residual phosphorus concentration (dissolved and particulate phosphorus) is not reduced to less than 0.3 mg/l until the pH approaches 11.5. The efficient clarification and phosphorus removal observed at pH values greater than 9 is related to the direct precipitation of the phosphorus and co-precipitation with the magnesium hardness as shown in Figure 6. Hence, magnesium hydroxide formation is desirable because the hydroxide acts as a flocculant and thus helps to clarify the water and to improve overall phosphorus removal. In contrast, in waste-waters with high alkalinities (∼350 mg/l $CaCO_3$), sufficient calcium
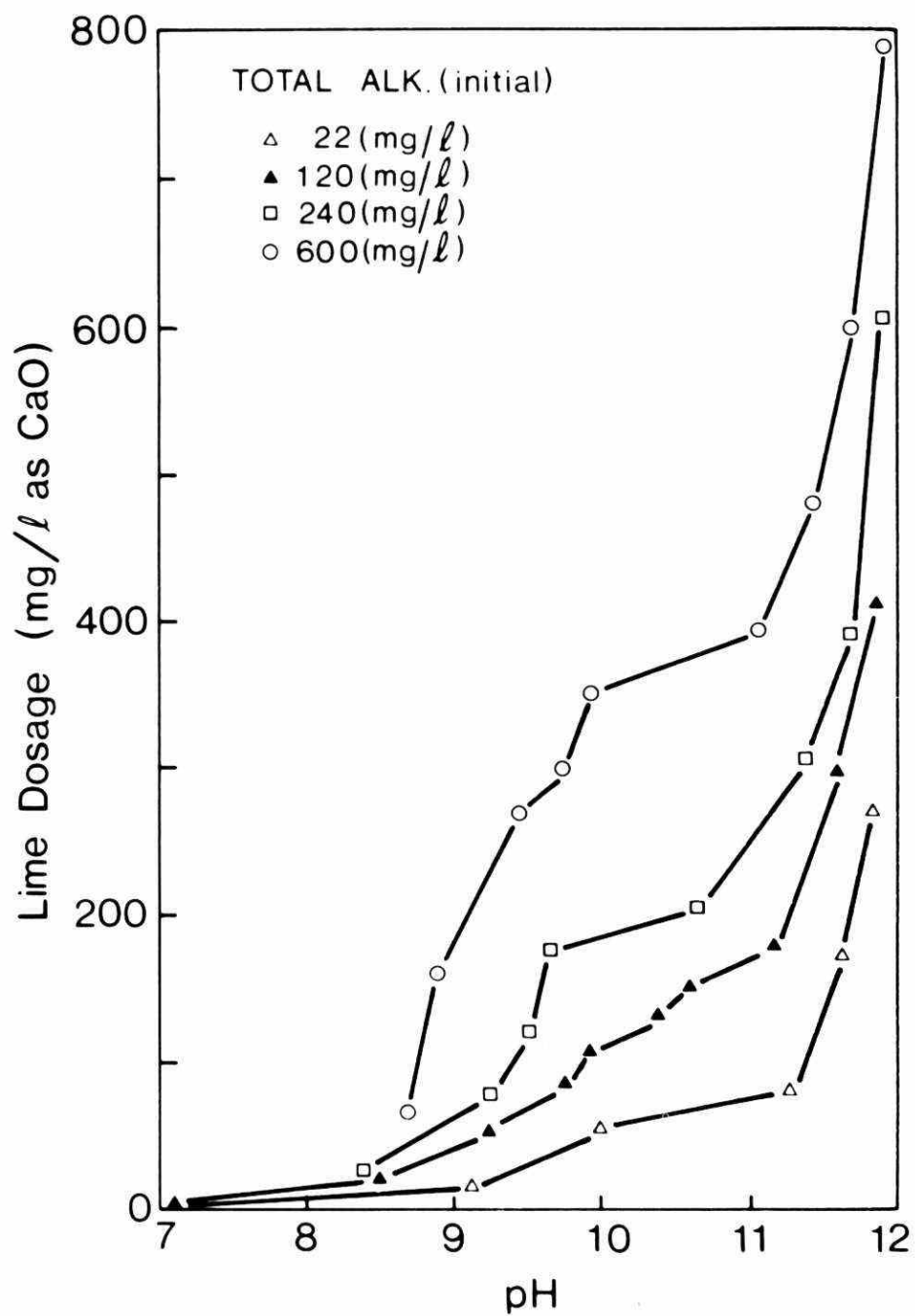
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FIG.4 RELATIONSHIP BETWEEN ALKALINITY, LIME
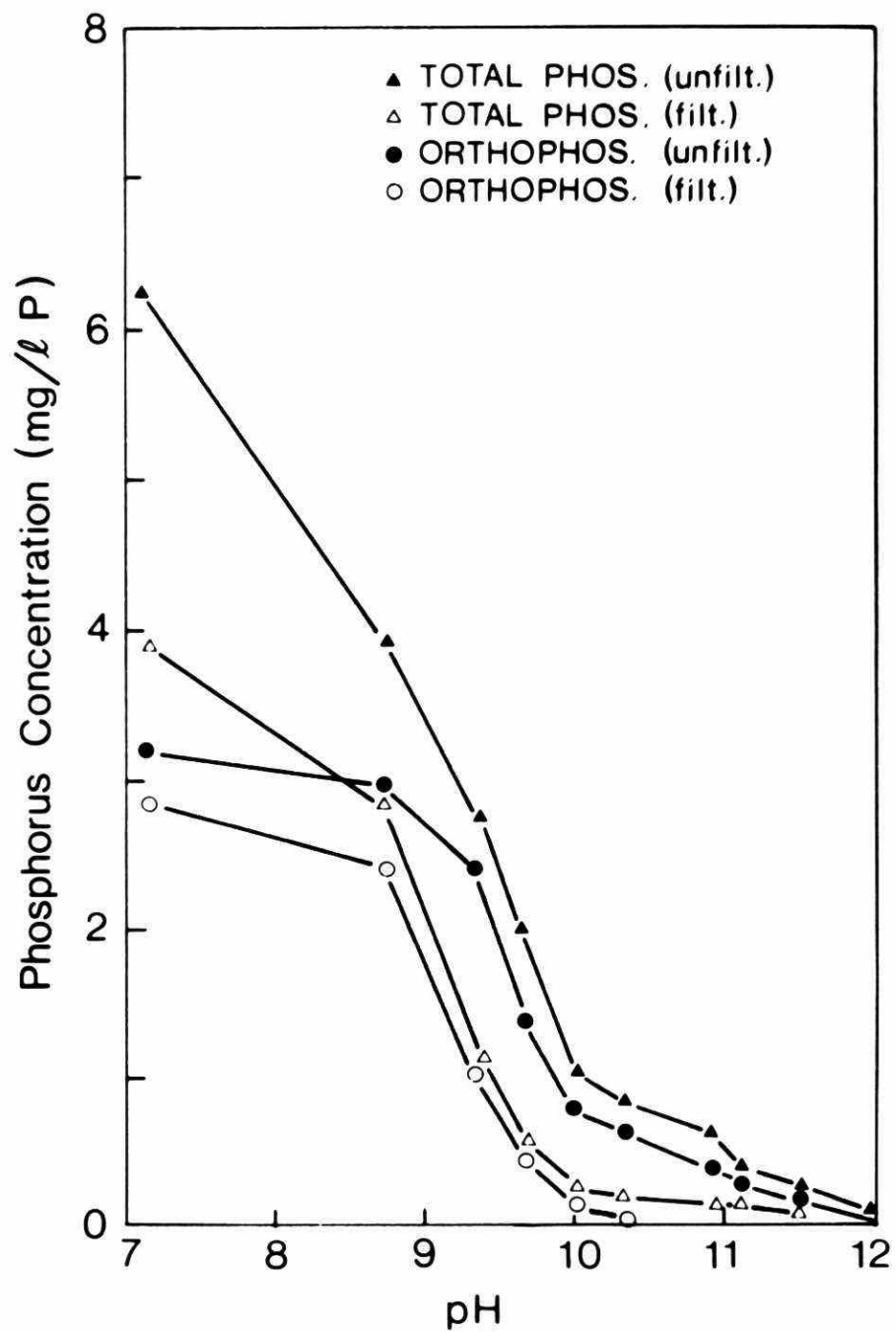DOSAGE AND pH (STAMBERG, ET AL., 1970)

FIG.5 PHOSPHORUS REMOVAL FROM RAW WASTE-
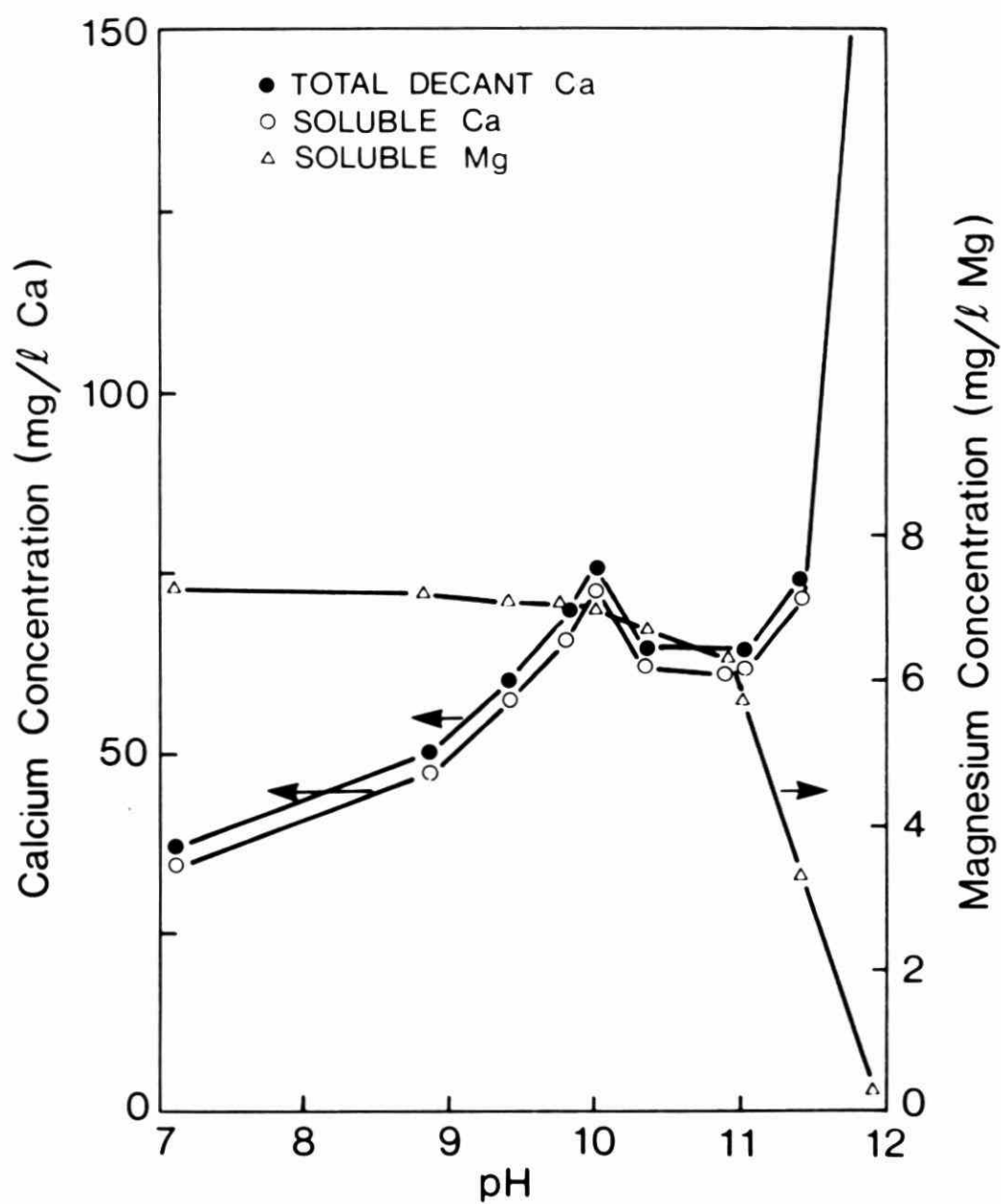WATER AS A FUNCTION OF pH
(STAMBERG , ET AL.,1970)

FIG.6 CALCIUM AND MAGNESIUM CONCENTRATIONS
IN RAW WASTEWATER AS A FUNCTION OF pH
(STAMBERG, ET AL.,1970)

carbonate precipitates in the pH range between 9.5 to 10 so as to produce
suspended solids that are filterable with a dual-media filter.

FULL SCALE DESIGN CONSIDERATIONS

Primary Treatment Plants

Phosphorus removal to an effluent total phosphorus concentration
of 1 mg/l has been successfully demonstrated at primary treatment plants
by the addition of lime, or aluminum or iron salts, to the raw wastewater.
Attaining effluent total phosphorus levels <0.3 mg/l will require
essentially complete insolubilization of soluble phosphates and efficient
capture of the phosphate solids, both inorganic and organic.

Theoretical considerations and previously discussed experi-
mental work indicate that, in the case of aluminum/iron salt addition for
phosphorus removal, it might be advantageous to lower the wastewater pH
more towards the optimum range for phosphate precipitation. Neglecting
other factors such as economics for the moment, then at the normal pH
levels of municipal sewage (6.8 to 7.8), it would appear that alum addition
should be favoured over iron addition. Adjusting the pH for optimum
phosphorus removal could be accomplished by the addition of excess coagulant
or acid. The method of choice will depend on the prevailing pH of the raw
wastewater and the desired operating pH. Supplementary acid addition
might be advisable for high pH wastewaters (pH $\geq$ 7.5) to avoid excessive
dosages of coagulant which could result in the formation of poorly settle-
able flocs not effectively removed by ordinary sedimenation or filtration
methods. To achieve consistently low effluent phosphorus concentrations
from a primary treatment plant through addition of aluminum/iron salts
might also require polymer addition - to increase solids flocculation and
settleability - followed by multi-media filtration of the primary effluent
to remove any carry-over phosphate floc.

Lime addition to raw sewage has proven to be a successful method
of obtaining low phosphorus concentrations in primary treatment plant
effluents. O'Farrell and Bishop (1971) reduced total phosphorus concen-
trations in a low alkalinity (142 mg/l $CaCO_3$) wastewater to <0.3 mg/l
with the addition of 400 mg/l $Ca(OH)_2$. Their pilot plant was operated
as a two-stage high lime process (pH = 11.5) and consisted of first-stage

chemical clarification, recarbonation, flocculation, second-stage clari-
fication and dual-media filtration. The pilot plant operating parameters
and effluent quality are summarized in Table 2. Clarified effluent
total phosphorus concentrations of 0.18 mg/l were obtained when ferric
chloride was used as a flocculating aid.

In general, the choice between a single-stage and a two-stage
lime process will depend on the alkalinity of the wastewater and the required
operating pH. In cases where chemical recovery is desired, the high lime
process (pH $\geq$ 11) would be required. Culp and Culp (1972) have indicated
that lime recovery (via recarbonation and calcination) is economically
feasible for treatment plants >10 MIGPD. However, an economically viable
process for aluminum or iron recovery has not been developed at this
time.

Chemical Addition to Secondary Treatment Plants

For the purpose of this paper, secondary treatment plants are
limited to those utilizing the activated sludge process or one of its
modifications (e.g., extended aeration, contact stabilization). As a rule,
in activated sludge plants alum or iron salts are added either to the
raw sewage ahead of the primary clarifier, or near the outlet of the
aeration tanks. Phosphorus removal by lime addition is not practicable
in the activated sludge process because of the high operating pH required
for effective phosphorus removal.

In general, equivalent phosphorus removals are obtained at
lower aluminum or iron dosages in a secondary treatment plant as compared
to a primary plant. This can be attributed to the biological uptake of
phosphorus by the activated sludge mass and the phosphorus removal capa-
bilities of the combined chemical/biological return activated sludge. As
a result, aluminum and iron dosages required to achieve effluent total
phosphorus levels of 0.3 mg/l will be less than those required in a
primary plant aiming at the same effluent quality objective. Table 3
lists some secondary treatment plants in Ontario that achieved effluent
total phosphorus concentrations close to or less than 0.3 mg/l with
iron or aluminum addition during plant scale treatability studies (Prested,
et al., 1975). Al or Fe to soluble phosphorus ratios necessary to achieve
these low effluent total phosphorus concentrations are approximately 1.3

TABLE 2   TWO-STAGE LIME TREATMENT OF RAW WASTEWATER
(O'FARRELL AND BISHOP, 1971)

A) Operating Parameters

Alkalinity (mg/l $CaCO_3$)	Lime Dosage (mg/l CaO)	1st Stage pH	2nd Stage pH	Flocculation Aid
142	300	11.5	10.1	5 mg/l Fe^{3+}

B) Pilot Plant Performance

	BOD		SS		Total Phosphorus	
	mg/l	% Removal	mg/l	% Removal	mg/l	% Removal
Influent	129	-	159	-	8.4	-
Clarified Effluent	24.3	81	18.5	88	0.27	97
Filtered Effluent	20.4	84	5.8	96	0.18	98

TABLE 3 ONTARIO TREATMENT PLANTS THAT ACHIEVED $\approx$0.3 mg/1 EFFLUENT
TOTAL PHOSPHORUS DURING PHOSPHORUS REMOVAL TREATABILITY STUDIES

A) Average Operating Conditions

Plant & Type	Influent [1]TP (mg/1)	[2]FP (mg/1)	Chemical Dosage	[3]Pt. of Addition	Sec. Clarifier SOR (Igpd/ft^2)
Bolton (Conv.Act.Sl.)	6.2	4.2	11 mg/1 Al	M	654
CFB Trenton (Conv.Act.Sl.)	3.0	1.8	5.4 mg/1 Al + 0.5 mg/1 A23 polymer	R	617
London Pottersburg (Conv.Act.Sl.)	3-5	-	25 mg/1 Al	R & M	520
Tillsonburg (Conv.Act.Sl.)	8.5	3.8	12.5 mg/1 Al	M	235
Westminster Twp. (Ext.Aer.)	9.5	2.0	9.0 mg/1 Fe	M	912
Woodstock (Conv.Act.Sl.)	8.0	1.3	17 mg/1 Fe	M	655

B) Effluent Quality

Plant	TP (mg/1)	BOD (mg/1)	SS (mg/1)	Turbidity (JTU)	Molar Ratio Me/TP	Me/FP
Bolton	0.34	16	21	-	2.1	3.0
Trenton	0.2	13	15	5	2.1	3.0
Pottersburg	<0.3	-	-	-	5-10	-
Tillsonburg	0.3	9	9	-	1.7	3.8
Westminster	0.31	10	10	-	1.0	2.5
Woodstock	0.2	11	12	-	1.2	7.2

[1] TP = Total Phosphorus   [2] FP = Filterable Phosphorus   [3] R = Raw Sewage, M = Mixed Liquor

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to 2.0 times higher than the ratios normally required (viz., 1.5 to 2.0) to achieve effluent total phosphorus concentrations of 1 mg/l. In general, these plants are well operated and produce a good quality effluent with BOD values $\leq$20 mg/l and low turbidities. Except for Westminster, secondary clarifier overflow rates were generally less than the critical clarifier overflow rate of 600 to 700 Igpd/ft^2 delineated by Boyko and Rupke (1973). The effect of secondary clarifier overflow rate on liquid-solid separation efficiency - as measured by the effluent suspended solids concentration - is shown in Figure 7. At final clarifier surface overflow rates exceeding approximately 600 to 700 Igpd/ft^2, a steady deterioration in effluent quality occurs.

Because of the relatively short duration (two weeks to one month) of the foregoing treatability studies, it is not known if these plants could sustain such low effluent total phosphorus concentrations on a long term basis. Although chemical/biological systems can achieve good conversion of soluble phosphorus to particulate phosphorus, filtration of the secondary effluent could be required to remove any carry-over floc and give consistent effluent total phosphorus concentrations of 0.3 mg/l or less. Generally, gravity and pressure filters of the multi-media type are used for the purpose of effluent polishing.

As in the case of primary treatment plants, chemical phosphorus removal in secondary treatment plants requires establishment and close control of all operating parameters. Hais, et al. (1971) investigated the use of dual and tri-media filters for polishing the effluent of an extended aeration plant using alum addition for phosphorus removal. Secondary effluent total phosphorus concentrations of 1.9 mg/l were reduced by tri-media filtration to 0.23 mg/l when alum addition reduced the secondary effluent pH to 6.5. Operation at pH levels >6.7 produced a fine phosphorus turbidity in the secondary effluent which was difficult to remove by filtration. Operation below pH 6.2 produced an upset in the biological system. They concluded that, with proper pH control (maintaining pH values in the range of 6.3 to 6.6) efficient phosphorus removal can be achieved in the activated sludge process with an Al:P weight ratio considerably less than 2:1 (Al:P molar ratio of 2.4:1). The relationship between filtered phosphorus residual concentrations
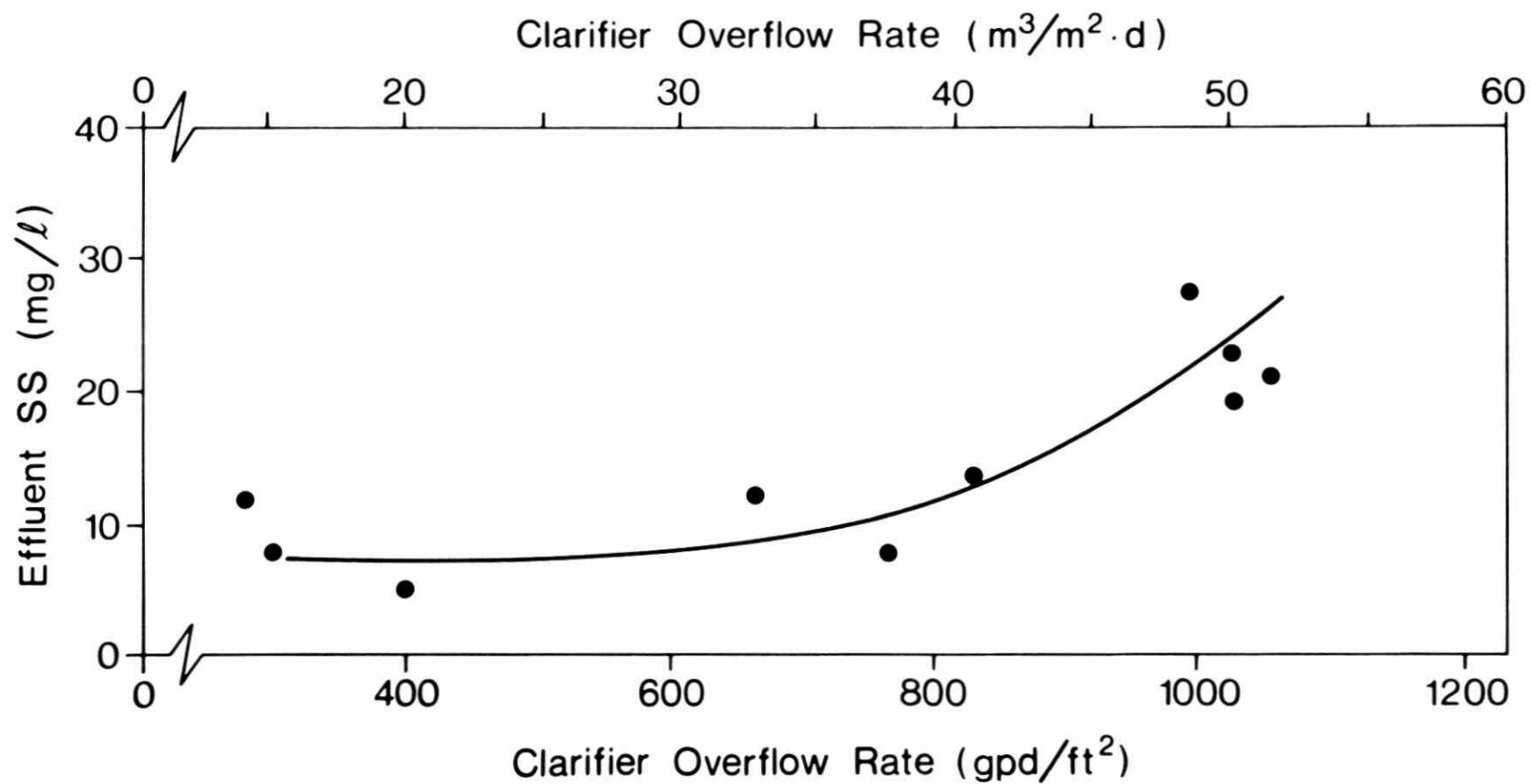
FIG.7   EFFECT OF FINAL CLARIFIER OVERFLOW RATE ON LIQUID-SOLID
SEPARATION (Aℓ^{3+} OR Fe^{3+}ADDITION TO AERATION SECTION)(BOYKO 1975)

and operating pH is shown in Figure 8. To control pH in the relatively narrow range of 6.3 to 6.6 would probably require alternate dosing of acid and base subject to the prevailing conditions of pH and alkalinity.

Post-Secondary Processes

Post-precipitation (chemical addition to the secondary effluent) is a well documented method of obtaining effluent total phosphorus concentrations in the range of 0.1 to 0.5 mg/l. Design considerations have been established (Culp and Culp, 1971; Shannon, 1973) and will not be discussed here to any great extent.

Post-precipitation requires the addition of separate equipment to an existing secondary plant. Individual components required are: flash mixing, flocculation and clarification (or alternatively a reactor clarifier combining all three), pH adjustment (if required) and filtration. Typical design parameters of the individual process components are summarized in Table 4. As in the case of secondary clarifiers, relatively low surface overflow rates are required for good solid-liquid phase separation.

In general, lime or alum are used as the chemical precipitants. Iron salts are seldom employed because of the colouring effect that the iron may have on the final effluent. The choice between lime and alum will depend on (i) the relative effectiveness of the two chemicals for phosphorus removal, (ii) the plant size, and whether chemical recovery is desired (lime recovery is feasible for plant sizes of 10 MIGPD while aluminum recovery is at present economically unattractive at any plant size), and (iii) whether sludge dewatering is practiced (tertiary lime sludges are easier to dewater than alum sludges).

Wastewater characteristics (alkalinity, hardness, pH) will influence the efficiency of the chemicals and the required chemical dosages in much the same manner as discussed previously for primary treatment plants. Alum dosages of 200 to 300 mg/l (9 to 14 mg/l as Al) are usually required to achieve effluent total phosphorus concentrations (after filtration) of ≤ 0.3 mg/l. Alternatively, lime dosages of 300 to 600 mg/l $Ca(OH)_2$ (depending on wastewater alkalinity and hardness) are required.

Table 5 summarizes the results of several studies that employed a lime or alum post-precipitation process. It is apparent from these data
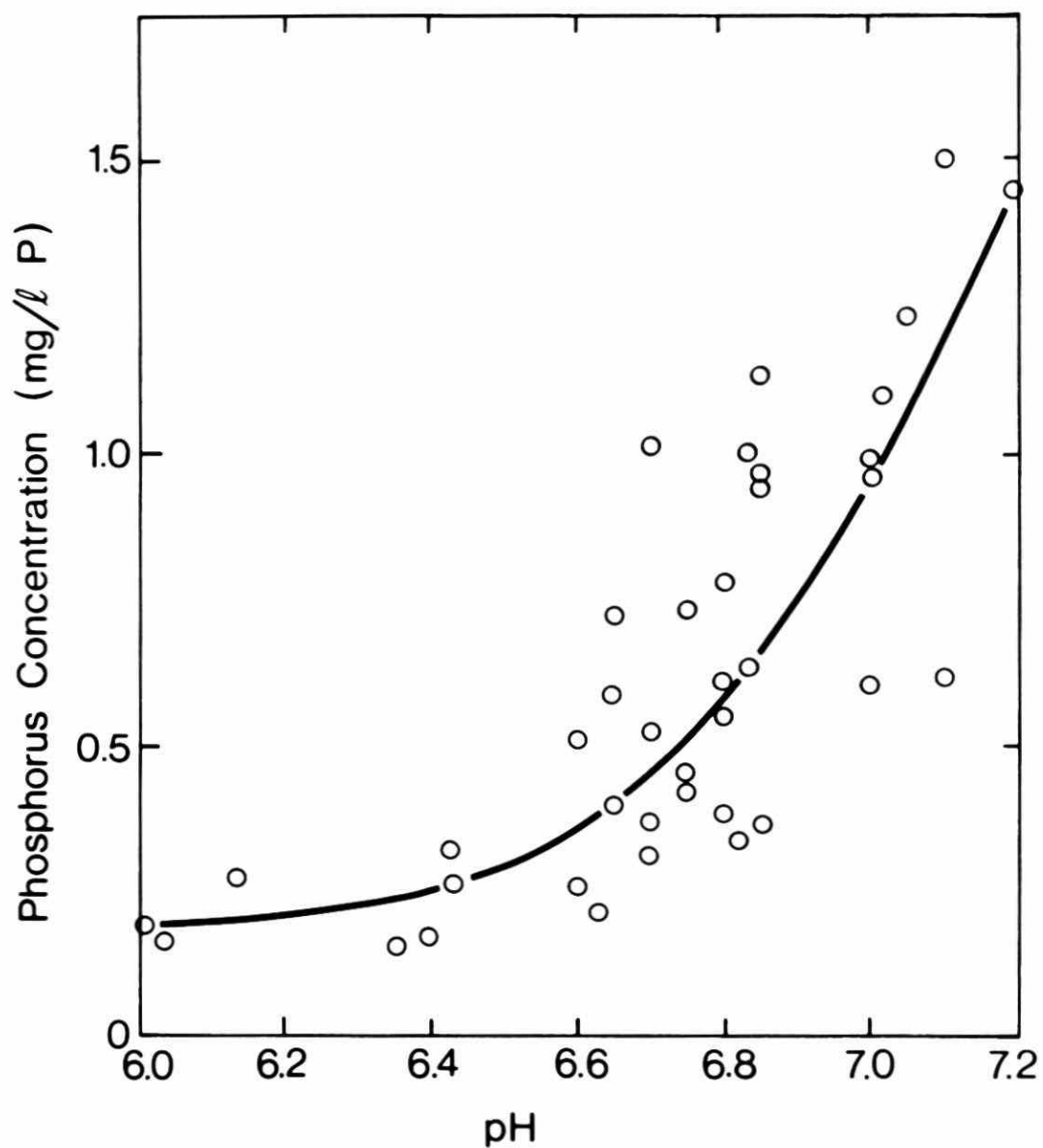
FIG.8 CORRELATION BETWEEN PROCESS pH AND PHOS-
PHORUS CONCENTRATION IN FOURTH - STAGE STEP
AERATION EFFLUENT (HAIS, ET AL.,1971)

TABLE 4   TYPICAL COMPONENT DESIGN PARAMETERS FOR A POST-PRECIPITATION
PHOSPHORUS REMOVAL PROCESS

	Process Component			
Chemical	Flash Mixing (Minutes)	Flocculation (Minutes)	Clarification SOR ($Igpd/ft^2$)	Filtration[1] Loading ($Igpm/ft^2$)
Lime	3 - 4	5 - 15	900 - 1400	$2.5^2 - 4.2^3$
Alum	1 - 2		700	

[1]   Mixed Media type most commonly used

[2]   Gravity Flow

[3]   Pressure Filtration

TABLE 5 POST-PRECIPITATION OF SECONDARY EFFLUENTS

Plant Location & Effluent Stream Treated	Process	Operating Conditions			Secondary Effluent		Tertiary Effluent	
		Chemical Dosage	1st Stage pH	2nd Stage pH	TP(mg/1)	Alkalinity*	Clarif. mg/1 P	Filt. mg/1 P
Ely, Minnesota; Sec. Trickling Filter Effluent	2-stage Reactor Clarifiers & Dual Media Filtration	Lime 150-200 mg/1 $Ca(OH)_2$	11.5	10.0	4-8	80-120	-	<0.2
Dist. of Columbia EPA-DC Pilot Plant; Sec. Step-Aeration Effluent	2-stage Reactor Clarifiers & Dual Media Filtration	Lime 460-610 mg/1 $Ca(OH)_2$	11.8	10.3	6-7	70-117	0.13	0.09
" "	1-stage Reactor Clarifiers & Dual Media Filtration	Lime 450-715 mg/1 $Ca(OH)_2$ & Na_2CO_3	11.6	-	6-7	80-146	0.53	0.10
Nassau County N.Y.; H.R. Act. Sludge Effluent	Single Stage Clarifiers & Dual Media Filtration	200 mg/1 Alum & polymer	-	-	9	-	-	0.10-1.0
Dayton, Ohio Pilot Plant; Act.Sl. Effl.	Single Stage Clarifiers & Dual Media Filtration	>200 mg/1 alum & polymer	-	-	-	-	-	0.3-0.6

* Alkalinity as mg/1 $CaCO_3$

TP - Total Phosphorus

200 mg/1 alum = 9 mg/1 Al

that, in general, lower and/or more consistent effluent phosphorus con-
centrations are obtained with the single-stage or two-stage lime process
than with alum addition.

COSTS FOR PHOSPHORUS REMOVAL

Costs associated with attaining effluent total phosphorus
concentrations less than 0.3 mg/l were estimated for the following
phosphorus removal schemes:

1) Aluminum or iron addition to the activated sludge
   process (Note: alum addition assumed for cost
   calculations);

2) Same as (1), plus multi-media filtration of the secondary
   effluent;

3) Post-secondary precipitation using lime.

In all cases it has been assumed that a secondary treatment
plant employing the activated sludge process already exists. Costs
have been limited to capital and operating costs of the phosphorus removal
system (chemical feed system) and any additional process equipment
(effluent filter, post-secondary precipitation equipment) that is required
to attain the phosphorus removal objective. However, it should be noted that,
in order to arrive at the optimum phosphorus removal scheme for a particular
wastewater treatment plant, sludge treatment and disposal costs would
also have to be considered.

Cost estimates for the above phosphorus removal process alter-
natives have been based on the following design assumptions:

a) Schemes 1 and 2 - chemical/biological

   - average total P = 6.0 mg/l and total filterable P =
     3.6 mg/l in influent wastewater
     (based on COA treatability study statistics)

   - chemical requirements: 15 mg/l Al, (based on an Al:FP
     molar ratio of 4:1)

b) Scheme 3 - Post-precipitation

   - a high pH ($\geq$11) single-stage lime process with effluent
     recarbonation

- lime requirement: 500 mg/l $Ca(OH)_2$

Total annual costs for the aforementioned phosphorus removal schemes are presented in Table 6 for plant flows of 1 and 10 MIGPD. The capital and operating costs (July 1972, ENR = 154.6) were obtained from Hazen and Sawyer (1975) and adjusted to September 1975 (ENR = 197). Capital costs have been amortized over 20 years at an interest rate of 10%.

Expressed on a basis of flow treated, the total costs range from a high of $216.45 per MIG for a 1 MIGPD plant using the single-stage lime post-secondary process to a low of $72.83 per MIG for a 10 MIGPD plant using a chemical/biological process with alum addition. As would be expected, the costs for phosphorus removal increase as the requirement for capital equipment increases. For the purpose of comparison, the estimated cost for phosphorus removal designed to meet an effluent total phosphorus objective of 1 mg/l (in the case of a 1 MIGPD plant adding 7.5 mg/l Al (Al:FP molar ratio of 2.0) to the mixed liquor) would be approximately $38.00. Thus, as a first estimate, the costs associated with achieving a total phosphorus concentration less than 0.3 mg/l in the effluent of an existing water pollution control plant using the activated sludge process could be two to five times the costs associated with achieving effluent total phosphorus concentrations of 1 mg/l.

It must be emphasized that the costs presented in Table 6 are primarily for illustrative purposes. They should only be used for a relative cost comparison of the three phosphorus removal schemes under the assumed conditions (phosphorus concentrations, chemical dosages, etc.). Estimates of phosphorus removal costs at any particular water pollution control plant should be based on treatability study data (jar test and/or full scale) for that particular treatment plant.

SUMMARY

1. To achieve effluent total phosphorus concentrations less than 0.3 mg/l requires essentially complete conversion of the soluble phosphorus to particulate phosphorus, followed by efficient removal of the phosphorus-containing solids. Once the chemical requirements are established for effective

TABLE 6   ESTIMATED TOTAL ANNUAL COSTS TO ATTAIN <0.3 mg/1 EFFLUENT TOTAL PHOSPHORUS*

Costs	Scheme 1[1]		Scheme 2[1]		Scheme 3[1]	
	1 MIGD	10 MIGD	1 MIGD	10 MIGD	1 MIGD	10 MIGD
Capital ($)	24,000.	84,000.	217,400.	917,500.	312,000.	1,287,500.
O & M ($)	25,600.	256,000.	41,600.	323,850	42,500.	306,000.
Amortized Capital @ 10% Interest ($)	2,808.	9,828.	25,436.	107,348.	36,504.	150,638.
Total Annual Costs ($)	28,408.	265,828.	67,036.	431,198.	79,004.	456,638.
Cost per Mil. Gal ($)	77.83	72.83	183.66	118.14	216.45	125.11

[1]   Scheme 1 - 15 mg/1 Al to activated sludge process

Scheme 2 - 15 mg/1 Al to activated sludge process and secondary effluent filtration

Scheme 3 - 500 mg/1 $Ca(OH)_2$ to single-stage post-secondary process

* See text for assumptions

conversion of soluble phosphorus to an insoluble form, the final effluent total phosphorus values are solely a function of solids capture in the treatment system. To obtain consistent effluent total phosphorus concentrations less than 0.3 mg/l will probably require tertiary effluent polishing via filtration.

2. Both soluble phosphorus conversion and solid-liquid separation processes are strongly dependent on process operating conditions. In the case of aluminum or iron salt addition, it could be advantageous to adjust the wastewater pH so as to optimize phosphorus precipitation and phosphorus removal processes. The method of adjusting the pH (excess coagulant and/or acid) is important, because excessive coagulant dosages can result in the formation of poorly settleable flocs which cannot be effectively removed by ordinary sedimentation or filtration methods. In the case of lime addition, it will be necessary to raise the wastewater pH to a level at which there is sufficient calcium and magnesium precipitation to effect good flocculation and efficient solid-liquid separation.

3. Because of the variability of wastewater characteristics and treatment plant facilities, the choice of the most advantageous phosphorus removal system for a particular treatment plant will have to be determined on a plant by plant basis. The relative effectiveness of the various chemicals commonly employed for phosphorus removal should be determined through treatability studies. In the case of alum or iron addition to the mixed liquor of a secondary treatment plant, more emphasis should be placed on plant scale studies rather than jar test studies, because jar tests do not reflect the enhanced phosphorus removal capabilities of the recycled chemical/biological sludge. Other factors that should be taken into consideration are: (i) the plant size, and whether or not chemical recovery is desired (feasible with lime for plants $\geq$10 MIGPD (Culp and Culp,

1971), not economically viable with alum); (ii) whether or not sludge dewatering is to be practiced (lime sludges are easier to dewater than alum sludges); and, (iii) sludge disposal practices (the increase in sludge volume is less with lime addition than with alum or iron salt addition; however, alum and iron sludges are amenable to anaerobic digestion, while operational problems may be encountered with anaerobic digestion of sludge containing lime).

4. Costs for phosphorus removal to effluent total phosphorus concentrations smaller than 0.3 mg/l will depend on the phosphorus removal system chosen (chemical/biological or post-secondary process). Capital costs increase with the requirement for additional equipment. Estimated costs for the conditions assumed in this paper range from $216.45 per MIG for a 1 MIGPD plant using the single-stage lime post-secondary process (500 mg/l $Ca(OH)_2$ addition to secondary effluent) to $72.83 per MIG for a 10 MIGPD plant using the chemical/biological process (15 mg/l Al addition to the aeration tanks). Any estimate of phosphorus removal costs for a particular wastewater treatment plant should, however, be developed on the basis of treatability study data as discussed herein.

REFERENCES

Black and Veatch, Consulting Engineers, "Process Design Manual for Phosphorus Removal". Report to U.S. Environmental Protection Agency, Program #170.0 GNP (1971).

Black and Veatch, Consulting Engineers, "Estimating Costs and Man-power Requirements for Conventional Wastewater Treatment Facilities". Report to U.S. Environmental Protection Agency, Project #17090 DAN (October 1971).

Boyko, B.I., "Phosphorus Removal Studies and Process Design Considerations". Presented at Technology Transfer Seminar on Alternatives for Nutrient Control; Environment Canada and B.C. Water Resources Service, Kelowna, B.C., Oct. 29-31, 1975.

Boyko, B.I. and Rupke, J.W.G., "Design Considerations in the Implementation of Ontario's Phosphorus Removal Programme". Phosphorus Removal Design Seminar; Environment Canada and Ontario Ministry of the Environment, Toronto, May 28-29, 1973.

"Canada/Ontario Agreement on the Lower Great Lakes." August 1971.

"Canada/United States International Agreement on Great Lakes Water Quality." April 1972.

Culp, R.L., and Culp, G.L. "Advanced Wastewater Treatment". Van Nostrand Reinhold Co., New York, N.Y. (1971).

Hais, A.B., et al., "Alum Addition to Activated Sludge with Tertiary Solids Removal". AIChE Symposium Series, 68 (124), 35 (1971).

Hazen and Sawyer, Engineers, "Process Design Manual for Suspended Solids Removal". Report to U.S. Environmental Protection Agency, EPA 625/1-75-003a (January 1975).

Jenkins, D., et al., "Chemical Processes for Phosphate Removal", Water Research, 5 (7), 369 (1971).

Lee, G.F., "Eutrophication". University of Wisconsin Water Resources Center (September 1970).

Marson, H., "The Removal of Phosphates from Sewage". Effluent and Water Treatment J., 11, 441 (1971).

Narkis, N., "Ferric Chloride or Lime for Chemical Treatment". Water & Sewage Works, 122(1), 43 (1975).

O'Farrell, T.P., and Bishop, D.F., "Lime Precipitation in Raw, Primary, and Secondary Wastewater". AIChE Symposium Series, 68(124), 43 (1971).

Prested, B.P., et al., "An Analysis of Canada/Ontario Phosphorus Removal Treatability Study Data". Canada/Ontario Agreement Research Report, Project 73-3-9, Part 1. In Press.

Recht, H.L., and Ghassemi, M., "Kinetics and Mechanism of Precipitation and Nature of the Precipitate Obtained in Phosphate Removal from Wastewater Using Aluminum (III) and Iron (III) Salts". U.S. Dept. of the Interior - Federal Water Quality Admin. (April 1970).

Schuessler, R.G., "Phosphorus Removal: A Controllable Process". AIChE Symposium Series, 67(107), 536 (1970).

Shannon, E.E., "Phosphorus Removal on Secondary Effluents". Phosphorus Removal Design Seminar; Environment Canada and Ontario Ministry of the Environment, Toronto, May 28-29, 1973.

Stamberg, J.B., et al., "Lime Precipitation in Municipal Wastewaters". AIChE Symposium Series, 67(107), 310 (1970).

Wilcox, R.L., "Removing in Excess of 99% Phosphorus at Ely, Minnesota". Papter presented at 73rd Meeting of Amer. Institute of Chemical Engineers, Minneapolis, Minnesota (August 1972).

DESIGN CONSIDERATIONS FOR BIOLOGICAL
CARBON REMOVAL-NITRIFICATION SYSTEMS

by

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INTRODUCTION

Active research in the past few years has led to a full awareness of the behaviour and effects of nitrogenous materials in aquatic environments. The deleterious effects of nitrogen in its various forms have led to a significant increase in basic and applied research aimed at the development of practical nitrogen control processes.

Ammonia removal or conversion is required for a number of reasons. The biochemical oxygen demand (BOD) of municipal wastes will be substantially influenced by the presence of ammonia. The conversion of ammonia to nitrate requires 4.6 parts of oxygen for each part of ammonium-nitrogen. Therefore, ammonia in the effluent at concentrations of 20 mg/l as N would give a theoretical nitrogenous oxygen demand of 92 mg/l.

Additional reasons for ammonia nitrogen removal include:
1) NH_3 at low concentrations is toxic to fish,
2) NH_3 is corrosive to copper fittings,
3) NH_3 increases Cl_2 breakpoint requirements and contact time for adequate disinfection, and
4)   eutrophication problems are associated with high nitrogen effluents.

Nitrogen control may involve either the nitrification of nitrogenous materials or total nitrogen removal. The latter can be accomplished by biological or physical-chemical means. The recent IAWPR conference proceedings (1975) entitled "Conference on Nitrogen as a

Water Pollutant" indicate the popularity of the biological process.

The process of nitrification and denitrification was recognized as occurring in wastewater treatment plants in the 1940's (Sawyer and Bradney, 1945; Brandon and Grindley, 1944). The simplicity of the process requirements and the absence of waste by-products have encouraged research studies in biological nitrification-denitrification. In the United States and Canada, the majority of domestic sewage treatment plants employ biological treatment, and biological nitrogen removal processes can take advantage of these existing facilities. Upgrading the treatment facilities to promote nitrification can eliminate the potential environmental problems resulting from the discharge of ammonia. If the resulting nitrate concentrations are not acceptable, denitrification can be added to the treatment process.

The efficiency of biological nitrogen removal is correlated to the degree of nitrification achieved. Autotrophic nitrifiers have lower yields and growth rates, and appear to be more sensitive to environmental conditions than the heterotrophic denitrifiers. Consequently, nitrification probably will prove to be the crucial step in the nitrification-denitrification process.

BIOLOGICAL NITRIFICATION

Principles

Biological oxidation of inorganic nitrogen is almost entirely carried out in the aquatic environment by bacteria belonging to the family Nitrobacteraceae. The major well defined autotrophic bacteria which are known to oxidize ammonia to nitrite are the genera Nitrosomonas and Nitrosococcus. The genera Nitrobacter and Nitrocystis have been shown to oxidize nitrite to nitrate.

The autotrophic nitrifying organisms derive all their energy requirements for growth and cellular metabolism from the free energy released by the oxidation of an inorganic nitrogen substrate. The source of carbon used for growth is obtained from carbon dioxide or bicarbonate. The bacterial oxidation reactions of ammonium to nitrite, and nitrite to nitrate, can be represented by the following equations:

$$NH_4^+ + 1.5\ O_2 \rightarrow 2\ H^+ + H_2O + NO_2^- \tag{1}$$
$$NO_2^- + 0.5\ O_2 \rightarrow NO_3^- \tag{2}$$
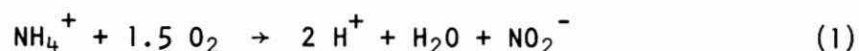
Process design requires rate data to be expressed as parameters which are useful from an engineering point of view. In order to describe the substrate removal rate in any biological waste treatment process, an overall kinetic expression compatible with the fundamentals of microbiology, kinetics, system hydraulics, and transport phenomena is necessary.

For a given reaction environment, substrate removal is a function of the concentration of substrate (S) and active biological solids (X):

$$-\frac{dS}{dt} = f\ (X,S) \tag{3}$$
$$-\frac{dS}{dt} = KXS^Z \tag{4}$$

where K is the substrate removal velocity. The unit rate of substrate removal can be expressed as:

$$-\frac{dS}{Xdt} = KS^Z \tag{5}$$

This expression has been used by numerous authors (Wuhrmann and Mechsner, 1965; Busch, 1971), and in fact, forms the basis for most engineering design work in biological treatment. The functional dependence of the substrate removal velocity depends on the kinetic model found applicable. Downing, Painter, and Knowles (1964) and Knowles, Downing, and Barrett (1965) used the Monod kinetic model in their rate expressions in analyzing nitrification results. Applying the expression to the ammonium oxidation by _Nitrosomonas_, the K_s values determined were small (0.2 to 1.7 mg/l NH_3-N).

At low K_s values, this model is very close to zero-order, independent of substrate concentration. The zero-order nature of the nitrification reaction, down to very low NH_3-N values, has been illustrated by numerous researchers (Wild, Sawyer, and McMahon, 1971;
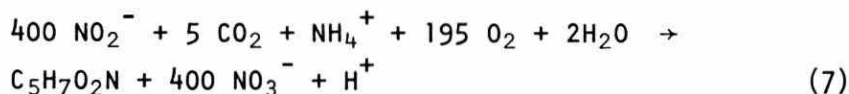
Kiff, 1972; Sutton, Murphy, and Jank, 1975). It is reasonable to
anticipate that the reaction rate will be zero-order with respect to
substrate concentration at all practical NH_3-N levels. This is con-
sistent with Monod kinetics at low K_s values. In such a situation the
unit rate of nitrification is equal to a constant, independent of
mixing regime.

Using reported values of actual cell yields and $C_5H_7O_2N$
as an empirical cell formula for nitrifying bacteria, Haug and McCarty
(1972) proposed the following overall mass balances for the nitrification
reactions:

Nitrosomonas

$$55 \ NH_4^+ + 5 \ CO_2 + 76 \ O_2 \rightarrow$$
$$C_5H_7O_2N + 54 \ NO_2^- + 52 \ H_2O + 109 \ H^+ \tag{6}$$

Nitrobacter

$$400 \ NO_2^- + 5 \ CO_2 + NH_4^+ + 195 \ O_2 + 2H_2O \rightarrow$$
$$C_5H_7O_2N + 400 \ NO_3^- + H^+ \tag{7}$$
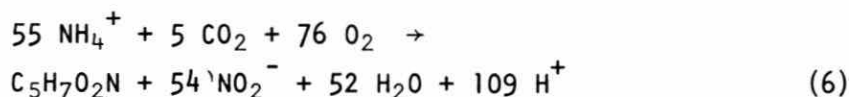
On the basis of the equations, 20 mg of ammonium-nitrogen would produce
only 3 mg of Nitrosomonas and approximately 0.5 mg of Nitrobacter.
These yields are less than 10 percent of that normally observed for
heterotrophic bacteria.

The growth of microorganisms may be expressed in terms of
their doubling or generation time. Generation times of the autotrophic
nitrifying bacteria are in the range of 10 to 30 hours. Generation
times for heterotrophic bacteria are frequently reported as 20 to 40
minutes. In a suspended growth or activated sludge system a direct
consequence of the slow growth rate or long generation time of nitrifiers,
is the requirement of a sufficient solids retention time (SRT) or
sludge age to retain an adequate population of these organisms.
Solids retention time is a measure of the average retention time of
the bacterial cells in the system. In an activated sludge system SRT
is normally defined as the total mixed liquor suspended solids under

aeration divided by the daily solids lost in the effluent or through sludge wasting.

In stating unit filterable total Kjeldahl nitrogen (TKN) removal rates for comparison of nitrification alternatives, it is necessary to state the operating solids retention time. Large differences in reported nitrification rates can often be attributed to the variation in the ratio of nitrifiers to the total microorganism population in the sludge. Activated sludge systems operating under similar conditions with equal solids retention time should contain a comparable fraction of nitrifiers.

Environmental Factors

Autotrophic nitrifying organisms are obligate aerobes. Numerous reports indicate that in order to ensure that dissolved oxygen is not a limiting factor in the nitrification reaction, a level not less than 2.0 mg/l must be maintained (Wuhrmann, 1963; Painter, 1975). The stoichiometric oxygen requirements based on equations 1 and 2 are 3.43 mg oxygen/mg ammonium-nitrogen and 1.14 mg oxygen/mg nitrite-nitrogen. Jeffrey and Morgan (1959) found that oxygen uptake values in BOD tests for nitrification were within 2.5 percent of the theoretical values.

Nitrification, like most bacterial processes, is affected by pH conditions. Generally, optimum conditions have been found to exist between pH 8.0 and 9.0 (Figure 1). Variations in pH optima could be due to shock effects in adjusting culture conditions or improper acclimation. Haug and McCarty (1972) using a submerged aerobic filter found that after an acclimation period of approximately 10 days the rate of nitrification at pH 6.0 was similar to rates within the normal range of operation, a pH of 7 to 8.5.

Nitrification is a hydrogen ion producing reaction (equation 1). Since pH values less than 6.0 would likely affect the nitrification rate the alkalinity of the waste is an important consideration. Assuming the pH is less than 8.5, the hydrogen ions produced during nitrification react with the bicarbonate in the wastewater, resulting in an increase in CO_2 concentration and a decrease in bicarbonate
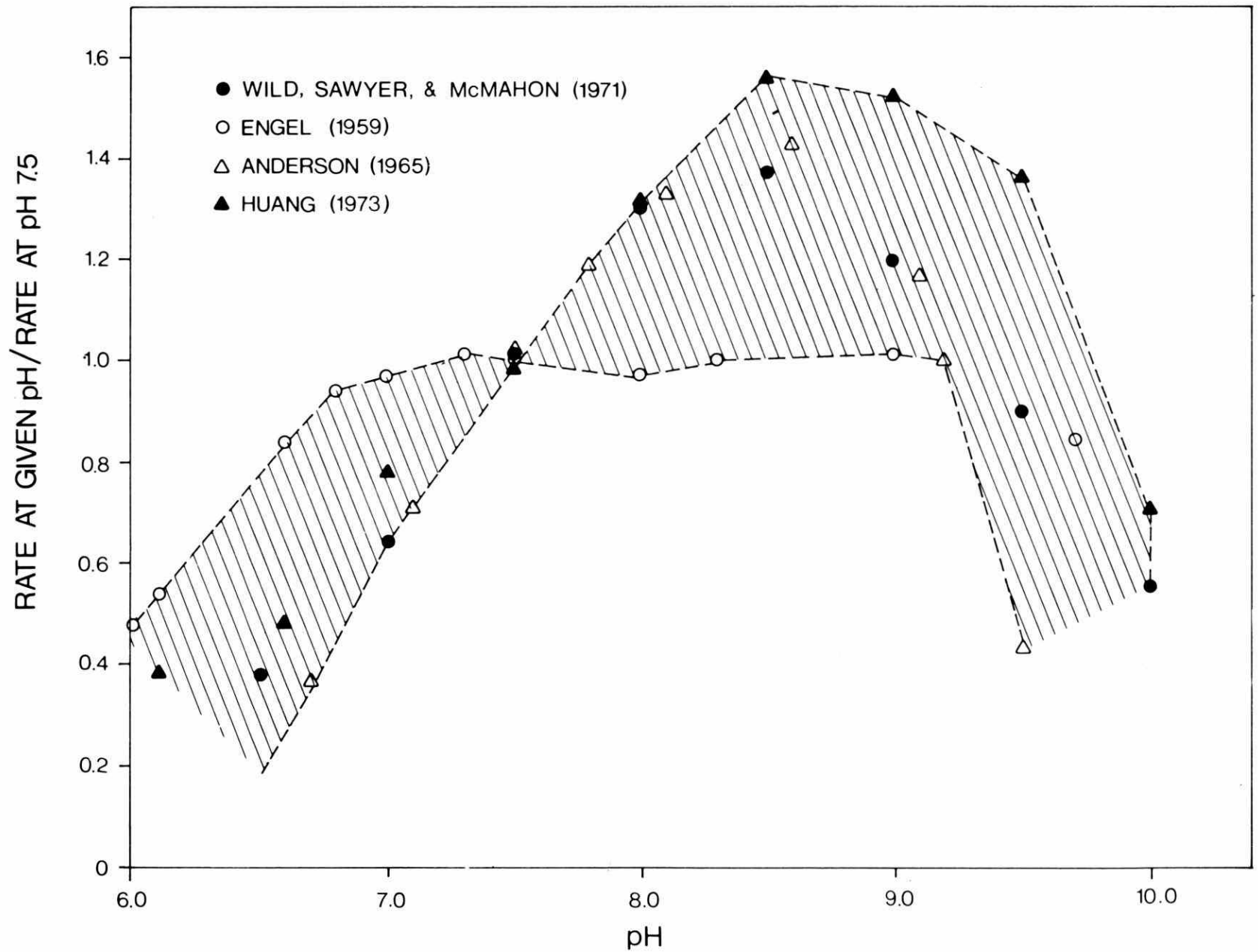
FIGURE 1. EFFECT OF pH ON NITRIFICATION RATE

alkalinity according to:

$$H^+ + HCO_3^- \rightarrow CO_2 + H_2O \qquad\qquad (8)$$

Based upon equations 1 and 8, approximately 7 mg of bicarbonate alkalinity, expressed as $CaCO_3$, are required to neutralize the hydrogen ions produced during the oxidation of 1 mg of ammonia-nitrogen. Calculations involving carbonic acid equilibria show that for wastewater with an alkalinity of 200 mg/l as $CaCO_3$, approximately 20 mg/l of NH_3-N could be oxidized before the pH dropped below 6.0 if all the CO_2 produced remained in solution (Haug and McCarty, 1972). In most nitrifying reactors the CO_2 is stripped from solution tending to help maintain a neutral pH. In operating a rotating biological contactor for carbon (BOD) removal-nitrification of a municipal sewage with moderate alkalinity (approximately 120 mg/l as $CaCO_3$), Wilson (1975) found that 15 to 20 mg/l NH_3-N were nitrified and the pH was never less than 6.8. Sutton (1975) observed similar results in operating suspended growth carbon removal-nitrification systems.

The process of nitrification, occurring over a range of approximately $4°$ to $45°C$ with optima at about $35°C$ for Nitrosomonas (Buswell, Shiota, Lawrence, and Meter, 1954) and $35°$ to $42°C$ for Nitrobacter (Deppe and Engel, 1960; Laudelout and van Tichelen, 1960) has been shown to be a strong function of temperature. In a suspended growth nitrification system differences in the reported temperature sensitivity may be due to differences in reactor SRT. Increasing the SRT results in a decrease in the temperature sensitivity. Supported growth systems (trickling filter, rotating biological contactor, etc.), which could be expected to be operating at high SRT's, show reduced temperature sensitivity for nitrification (Wilson, Murphy, Sutton, and Jank, 1975; Huang, and Hopson, 1974).

Nitrifying organisms, especially Nitrosomonas, are susceptible to a number of inhibitors which may be present in municipal and industrial wastewaters. A number of metals are toxic to nitrifiers but the concentration required to cause inhibition is dependent on the state of the culture. Copper, silver, mercury, nickel, chromium, and zinc

all may inhibit nitrification under certain conditions. Downing,
Tomlinson, and Truesdale (1964) identified a large number of organic
materials which caused a reduction in nitrification rates in suspended
growth systems. Thiourea, phenol, phenolic compounds, cresol, and
halogenated solvents were some of the potential inhibitors identified.
In screening the organic materials as nitrification inhibitors,
unacclimated organisms were used and, therefore, the results may not
reflect what actually would occur in a treatment plant.

Nitrification Treatment Alternatives

The process reactors available for carbon removal (BOD) -
nitrification can be classified according to the nature of their
biological growth. Activated sludge systems can be regarded as sus-
pended growth reactors whereas systems in which growth occurs on, or
within a solid media, can be termed supported growth reactors.

The two basic process schemes available for nitrification in
the activated sludge process are the combined sludge process and the
separate sludge process. In the combined sludge process, carbon
removal and nitrification are carried out using the same sludge. In
the separate sludge process, the biological reactions are carried out
by different microorganisms in separate reactors. In studying combined
sludge carbon removal-nitrification, Sutton, Murphy, Jank, and Monaghan
(1975) considered two alternatives; single and two-stage activated
sludge systems (Figure 2). The single-stage system could be approxi-
mated as a completely mixed reactor. The extent of mixing in the two-
stage reactor approached that normally found in conventional activated
sludge plants. Other combined sludge systems which can be designed to
incorporate partial or complete nitrification include: contact stabili-
zation, extended aeration, step aeration and oxygen activated sludge
plants (Brown and Caldwell, 1975). Combined sludge systems for
carbon removal-nitrification have proven to be a viable alternative to
the separate sludge system (Sutton, Murphy, and Jank, 1975; Lawrence
and Brown, 1973).

A limited number of supported growth systems have been
utilized for carbon removal and/or nitrification including trickling

TWO-STAGE SEPARATE SLUDGE
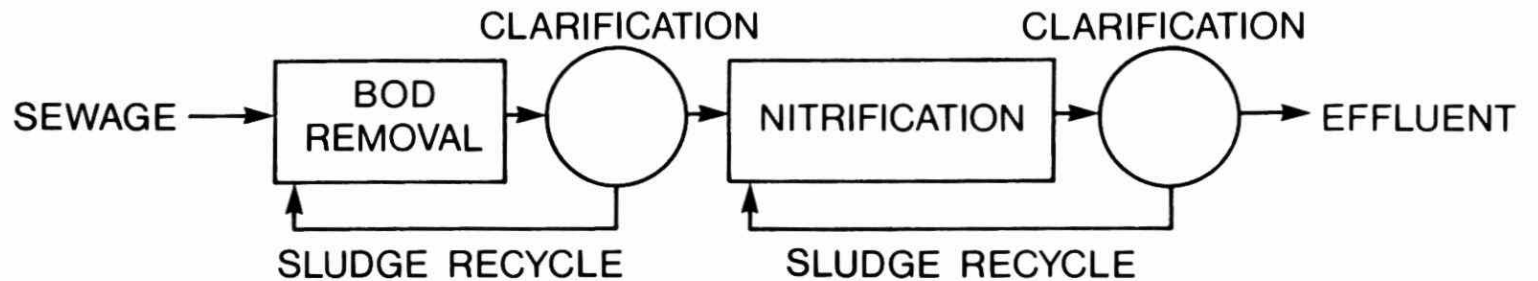
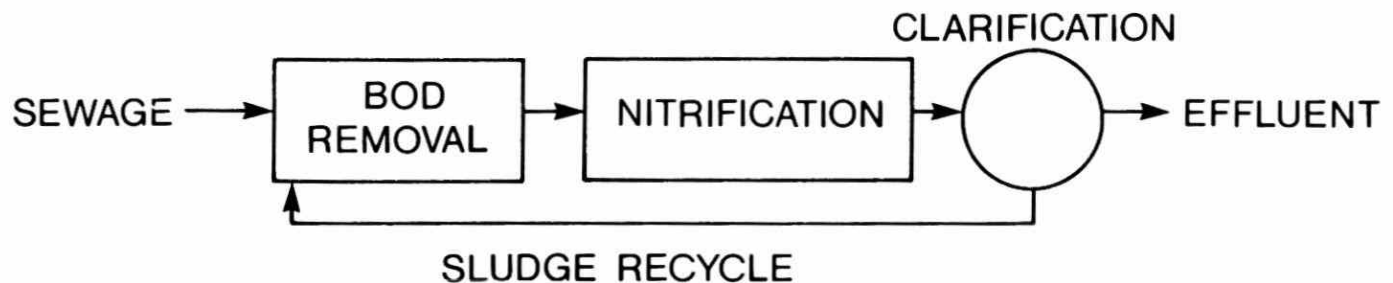SINGLE-STAGE COMBINED SLUDGE
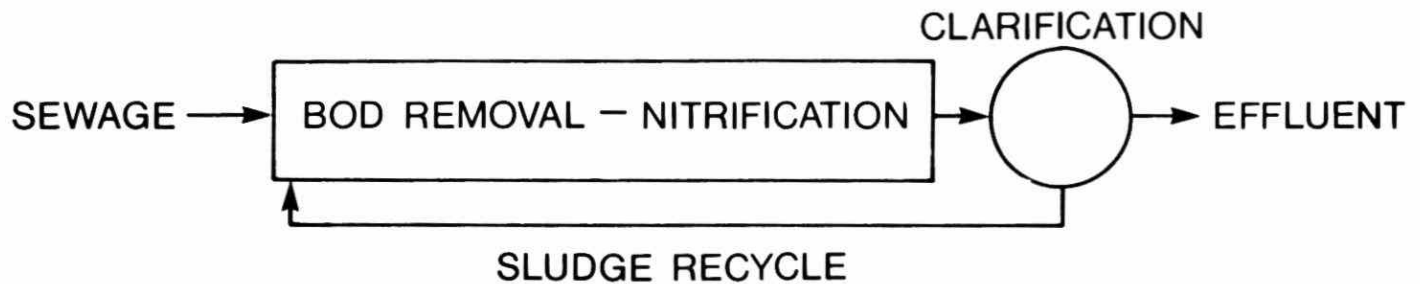
TWO-STAGE COMBINED SLUDGE

FIGURE 2. SEPARATE AND COMBINED SLUDGE CARBON REMOVAL —
NITRIFICATION SYSTEMS

filters, submerged aerobic filters, and rotating biological contactors (Figure 3).

Rocks, wood slats, and plastic sheets have all been used as media for trickling filters. In submerged upflow packed bed reactors, stones, gravel, plastic and other types of media can be employed. Although these packed tower reactors can be designed economically for nitrification down to ammonia-nitrogen levels approaching 2.0 mg/l, removals below this are difficult to attain (Stenquist, Parker, and Dosh, 1974).

The rotating biological contactor (RBC) consists of a series of closely spaced discs anchored to a shaft and partially submerged in the wastewater. The rotation of the shaft alternately exposes the biological slime growing on the disc surfaces to the waste and then to the atmosphere. Under proper loading conditions successful carbon removal-nitrification can be obtained (Antonie, 1974; Wilson, Murphy, Sutton, and Jank, 1975).

In addition to the nitrification process alternatives, a number of suspended and supported growth systems are available for denitrification.

PILOT PLANT PROGRAM

The many biological nitrogen conversion or removal schemes necessitate defining the important process parameters in order to allow a rational design selection. With this in mind, in May 1972, an intensive biological nitrification-denitrification research program began at the Wastewater Technology Centre, Burlington, Ontario, Canada. The pilot plant program, conducted as a joint project between McMaster University and the Environmental Protection Service, was geared to providing the required engineering data so that biological nitrification and denitrification systems could be designed and operated successfully.

To obtain the necessary design parameters for carbon removal-nitrification a 28.5 m^3/day (6280 Igpd) suspended and supported growth pilot plant was operated (Figure 4).
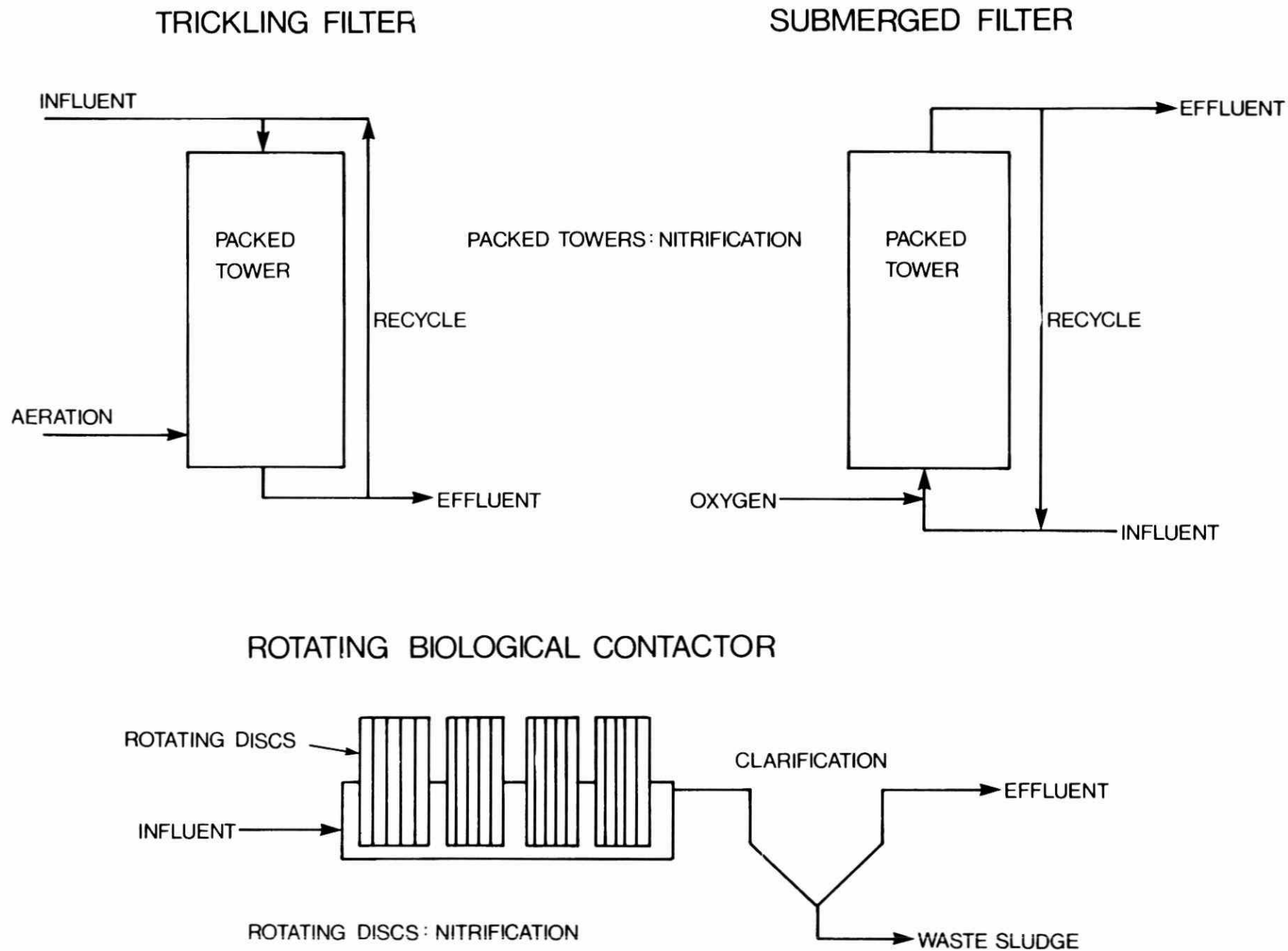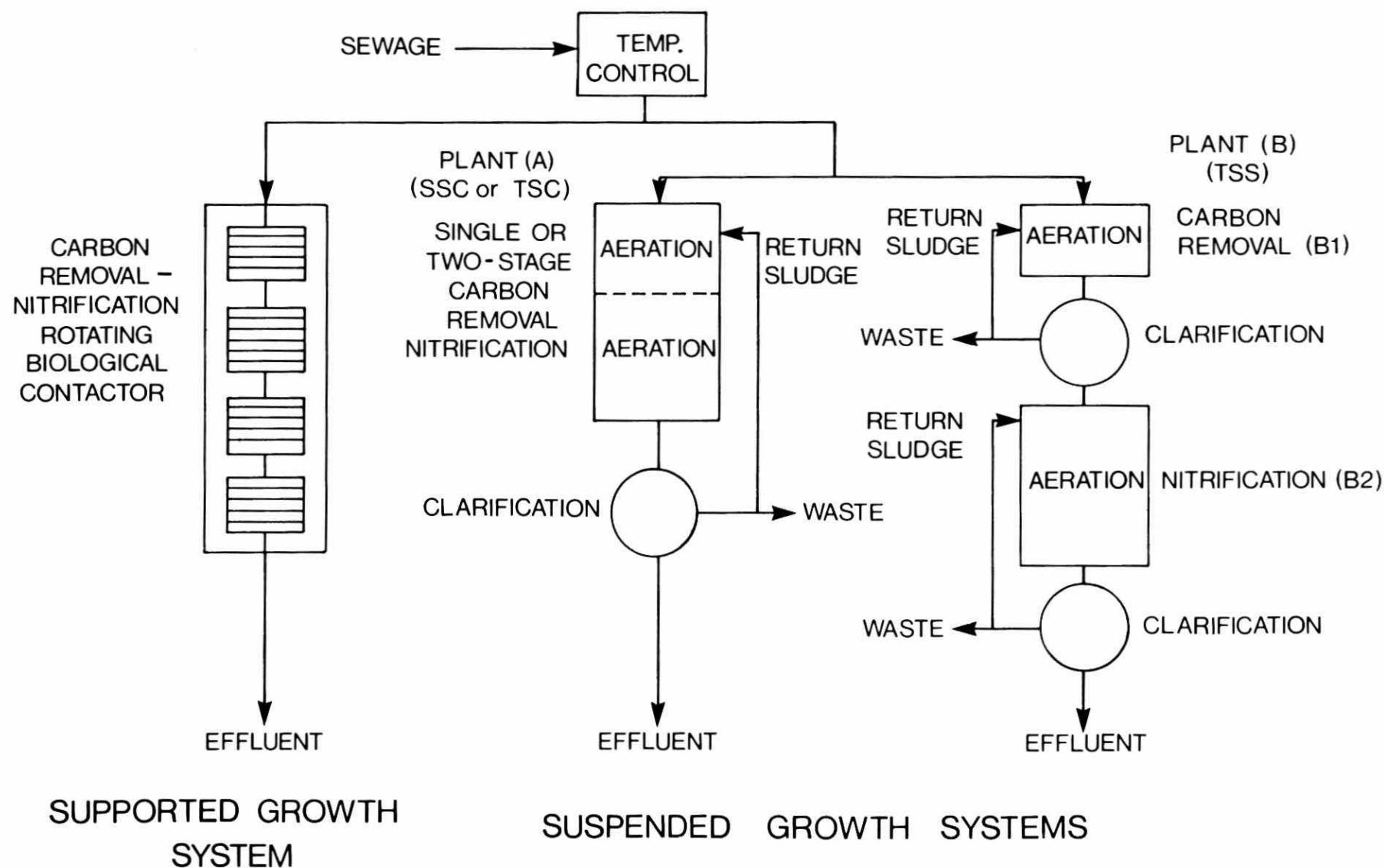
FIGURE 3. SUPPORTED GROWTH NITRIFICATION SYSTEMS

SEWAGE → TEMP. CONTROL

PLANT (A) (SSC or TSC)

PLANT (B) (TSS)

CARBON REMOVAL – NITRIFICATION ROTATING BIOLOGICAL CONTACTOR

SINGLE OR TWO-STAGE CARBON REMOVAL NITRIFICATION

AERATION

AERATION

RETURN SLUDGE

CLARIFICATION → WASTE

EFFLUENT

RETURN SLUDGE → AERATION   CARBON REMOVAL (B1)

WASTE ← CLARIFICATION

RETURN SLUDGE → AERATION   NITRIFICATION (B2)

WASTE ← CLARIFICATION

EFFLUENT

EFFLUENT

SUPPORTED GROWTH SYSTEM

SUSPENDED GROWTH SYSTEMS

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FIGURE 4. SUSPENDED AND SUPPORTED GROWTH CARBON REMOVAL–NITRI–FICATION PILOT PLANT

The suspended growth systems for carbon removal-nitrification, designated as plants A and B, consisted of two plexiglass, dispersed air aeration tanks. Plant A contained a removable divider allowing either the two-stage or the single-stage combined sludge system to be operated in parallel with Plant B. Plant B formed the two-stage separate sludge system, a fixed divider separating the reactor into two aeration tanks, B1 and B2.

The partially submerged rotating biological contactor operating in parallel with the suspended growth systems contained 36 discs with a total surface area of 23.2 m^2 (250 ft^2), a disc length of 1.32 m (4.33 ft) and a hydraulic capacity of 128 litres (28 Igal).

Further details concerning the reactor systems are presented elsewhere (Sutton and Jank, 1975).

The pilot plant reactors received degritted raw municipal wastewater subject to normal variation in organic and inorganic concentrations (Table 1). In order to determine the kinetic removal rates used to describe the performance of the nitrification systems, the effluent from the reactors must contain residual filterable ammonium-nitrogen. When complete ammonium conversion was obtained, additional amounts of ammonium (NH_4Cl) were added. Bicarbonate ($NaHCO_3$) was added to the feed when necessary to avoid any alkalinity limitation on the ammonium removal rates.

TABLE 1. RAW WASTEWATER CHARACTERISTICS

Characteristic (mg/l)	Mean	90%*
COD	325	460
BOD_5	120	200
SS	240	450
Filterable NH_4-N	15	18
Filterable TKN	17	25
Alkalinity (as $CaCO_3$)	118	130

*90% of observations are equal to or less than stated value

SUSPENDED GROWTH SYSTEM

Nitrification occurs in the activated sludge process when conditions are suitable for the retention and accumulation of nitrifying bacteria. Successful nitrification depends on adherence to a sludge wasting program which results in a SRT adequate to retain and prevent the washout of the slower growing nitrifying bacteria. The two basic process schemes available can be designated as:

1) a combined sludge system in which simultaneous carbon removal and nitrification are carried out, and

2) a separate sludge system in which carbon removal and nitrification are carried out by different microorganisms in separate reactors.

The combined sludge system may be single or multi-stage, while the separate sludge system is normally a two-stage system. Provided the rate of growth of nitrifying bacteria is rapid enough to compensate for the organisms lost through sludge wasting, nitrification can be maintained.

In this study carbon removal and nitrification efficiencies were obtained from three different process configurations, a single and two-stage combined sludge system (denoted SSC and TSC respectively), and a two-stage separate sludge system (TSS). Using 24-hour time-averaged sampling and parallel operation, the performance of any two alternatives were assessed. The alternatives were compared as to their effectiveness, in terms of nitrification, under a range of operating temperatures and solids retention times. By control of the SRT and temperature, a pseudo "steady-state" (steady flow and diurnal influent variations in organic and inorganic concentration) operation point was approached.

The three carbon removal-nitrification systems were compared first in terms of their ability to remove filterable TKN. The results from the analysis of paired data over a system sludge age range of four to ten days and temperature range of 5° to $25^\circ C$ using a "t-test" indicated comparable removal of filterable TKN by the separate (TSS) and combined sludge systems (TSC and SSC) at equal system solids

retention times. While the data might appear to favour the separate sludge system (Figure 5), a "t" value less than the critical value (95% level) for the differences (Di values) in TKN removal is indicated in Table 2. A direct comparison between the combined sludge systems (TSC and SSC) was not possible. Therefore, the mean differences ($\bar{D}$) in treatment observed between the SSC and TSS sludge system, and TSC and TSS sludge system, were examined statistically. The "t" value determined (Table 3) was not significant, implying that equal filterable TKN removal was obtained with both SSC and TSC sludge systems.

This comparison of the combined sludge reactors allowed an evaluation of the effect of reactor configuration on nitrification. Tracer analyses indicated that the single-stage reactor (SSC) could be approximated as a completely mixed system whereas the two-stage system (TSC), equivalent to two mixed tanks in series, was closer to conditions characteristic of full scale basins with large length to width ratios. The lack of difference in filterable TKN removal in the two combined sludge systems (Table 3), with substantially different mixing regimes, supports the "zero-order" nature of the nitrification reaction to very low filterable TKN or NH_3-N values (Wild, Sawyer, and McMahon, 1971; Downing, Painter, and Knowles, 1964).

TABLE 2.   DIFFERENCE IN TKN REMOVAL IN SEPARATE AND COMBINED SLUDGE SYSTEMS

Paired Data	$\bar{D}$	$\dfrac{S_D^2}{n}$	t	$\begin{array}{c} t \\ \alpha = .95 \end{array}$
TSS vs SSC	0.60	0.57	0.79	1.74
TSS vs TSC	1.16	0.44	1.52	1.72

Note: $\dfrac{S_D^2}{n}$ = estimated variance of $\bar{D}$, where

S_D^2 = $\dfrac{\Sigma\ (Di-\bar{D})^2}{n-1}$, and

n = number of paired data points

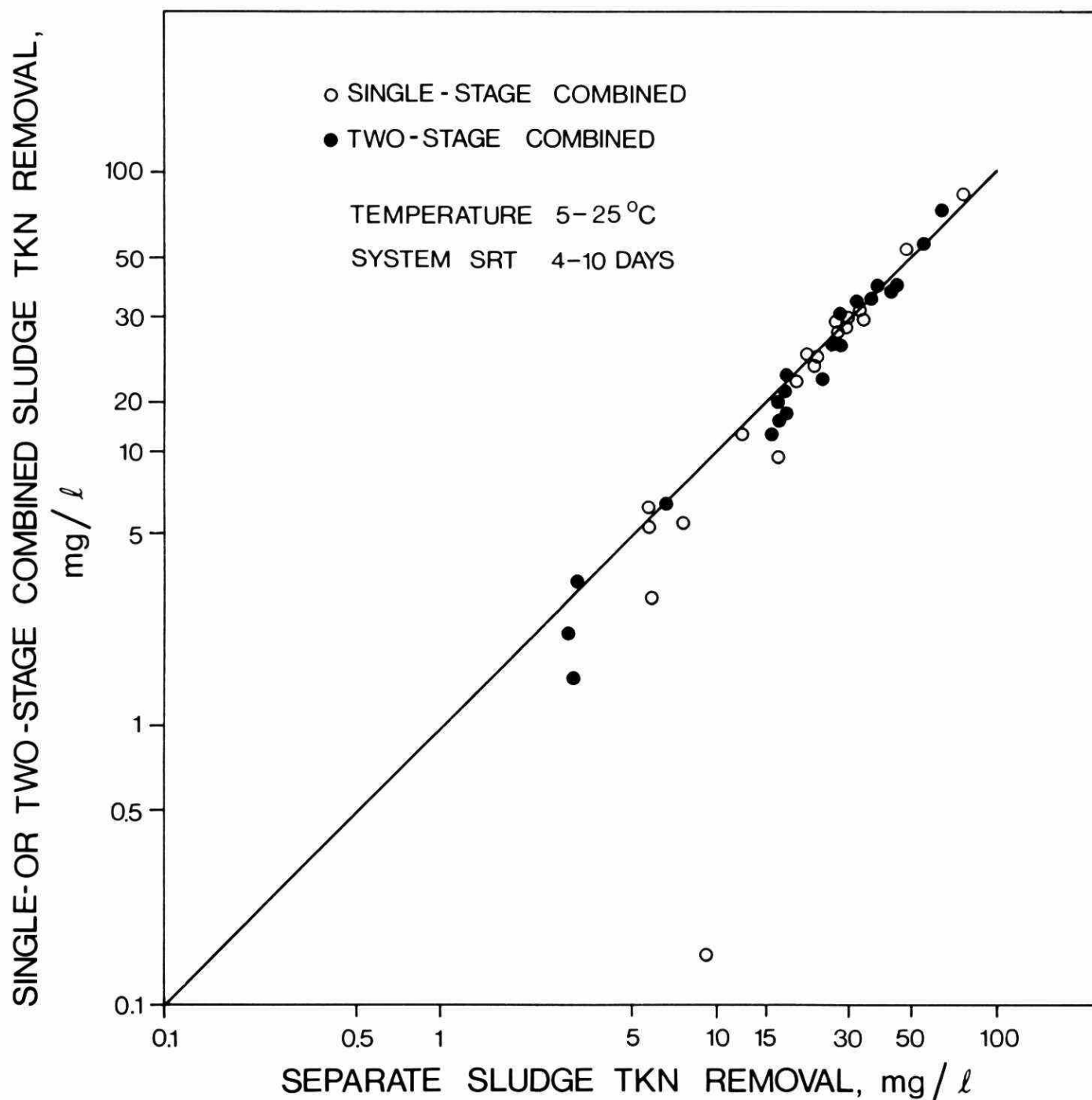t = $\dfrac{\bar{D}}{S_D/(n)^{\frac{1}{2}}}$

FIGURE 5. FILTERABLE TKN REMOVAL IN PARALLEL SEPARATE
AND COMBINED SLUDGE SYSTEMS

A further analysis of paired data suggested that the combined sludge system produced a better settling sludge (Figure 6). Statistical verification suggested that a real difference may not exist (Sutton, Murphy, and Jank, 1975). The mean clarifier effluent suspended solids concentration from both systems was 24 mg/l (Figure 7). This is typical of suspended growth systems of long SRT's.

TABLE 3. DIFFERENCE IN TKN REMOVAL IN SINGLE-STAGE AND TWO-STAGE COMBINED SLUDGE SYSTEMS

System Compared to TSS	$\bar{D}$	S_D^2	t	t $\alpha = .95$
SSC	0.6	10.82		
			0.56	1.68
TSC	1.16	9.67		

Note: $S_D^2 = \dfrac{\Sigma\,(D_i - \bar{D})^2}{n-1}$

n = number of paired data points

$t = \dfrac{\bar{D}_{TSC} - \bar{D}_{SSC}}{S_p\,(\dfrac{1}{n_{TSC}} + \dfrac{1}{n_{SSC}})^{\frac{1}{2}}}$, where

S_p^2 = pooled variance for $D_{TSC} - D_{SSC}$

$= \dfrac{n_{TSC}\,(S_{D\ TSC}^2) + n_{SSC}\,(S_{D\ SSC}^2)}{n_{TSC} + n_{SSC}}$

In certain instances, nitrification without subsequent denitrification may be required. In such cases, it is likely that a TKN limit will be specified. Even with complete nitrification, the clarified effluent can be expected to contain a small quantity of filterable organic nitrogen. This is probably associated with metabolic by-products. Current studies indicate filterable organic nitrogen concentrations averaging 0.8 to 0.9 mg/l. In addition, the clarified
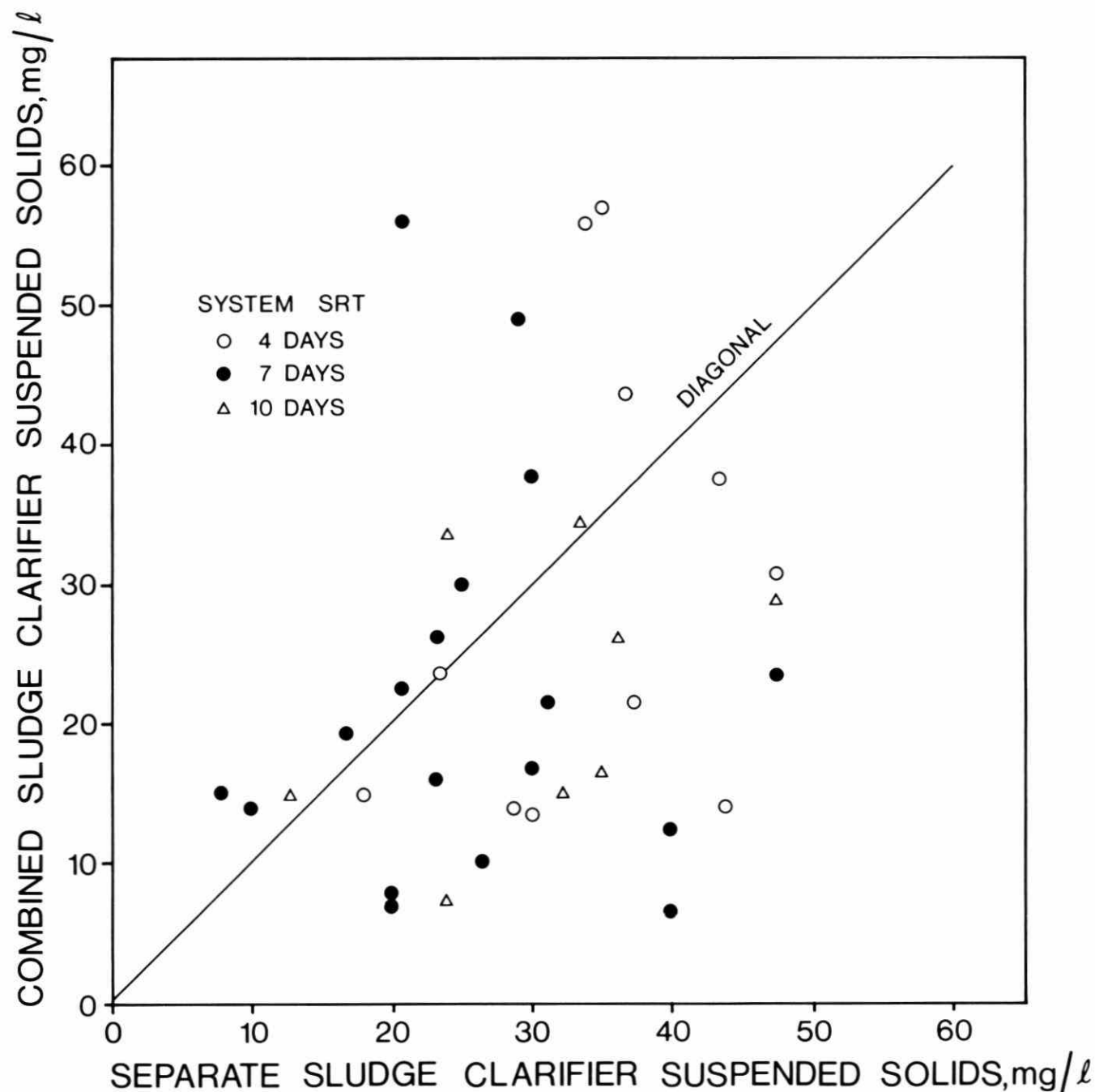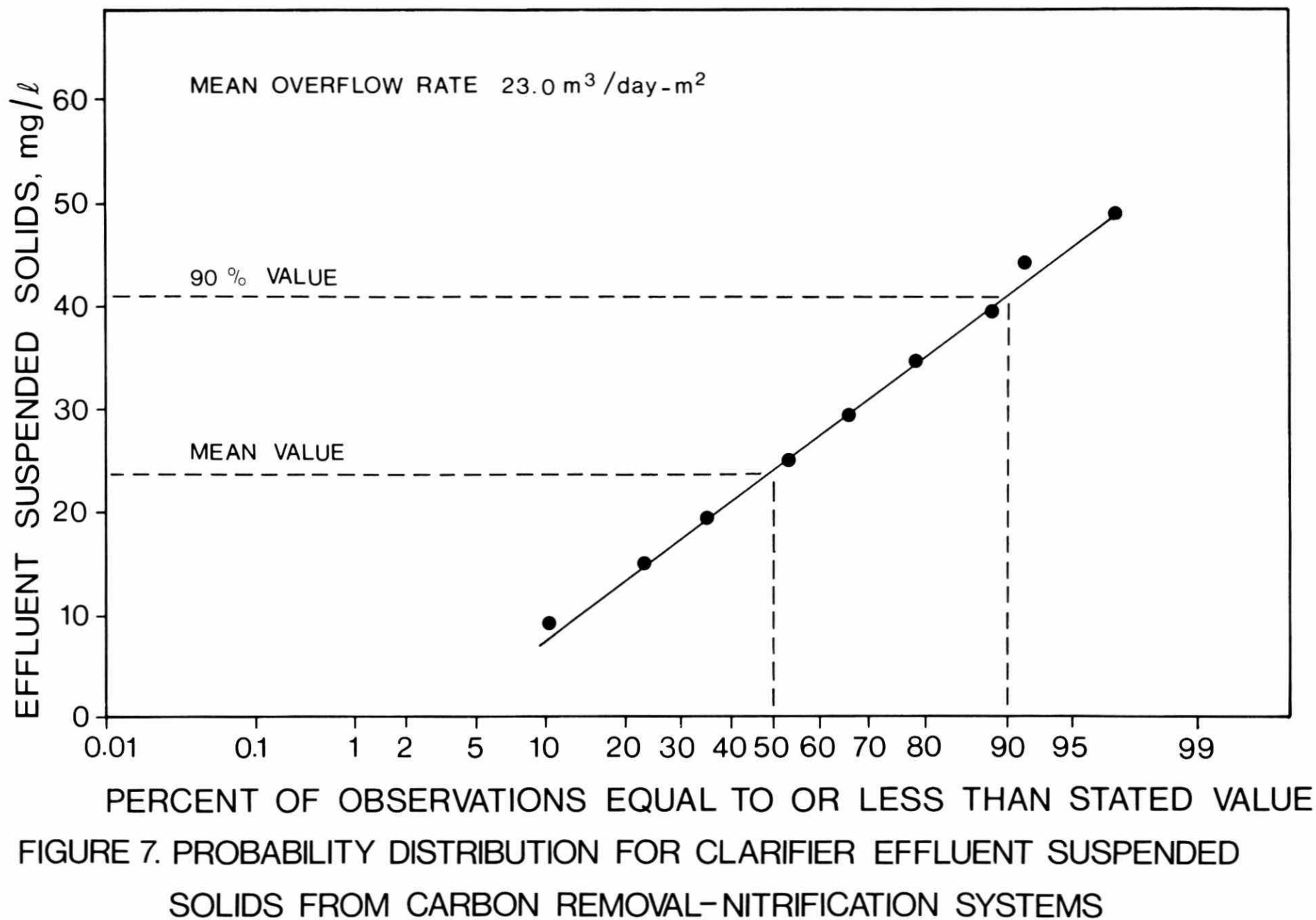
FIGURE 6. EFFLUENT SUSPENDED SOLIDS FROM PARALLEL
SEPARATE AND COMBINED SLUDGE SYSTEMS

FIGURE 7. PROBABILITY DISTRIBUTION FOR CLARIFIER EFFLUENT SUSPENDED SOLIDS FROM CARBON REMOVAL-NITRIFICATION SYSTEMS

effluent will contain 1.0 mg/l non-filterable TKN (Figure 8) caused by the presence of suspended solids as indicated in Figure 7.

The growth rate of <u>Nitrosomonas</u> and consequent rate of NH_3-N utilization has been reported to be a strong function of temperature (Knowles, Downing, and Barrett, 1965; Wild, Sawyer, and McMahon, 1971). For many reactions, the variation of rate with temperature may be represented by an Arrhenius relationship. This relationship can be written as:

$$K = Ae^{-E/RT} \tag{9}$$

where

K = reaction rate constant (hr^{-1})

A = frequency factor

E = activation energy (cal $g\text{-mole}^{-1}$)

R = universal gas constant (cal $g\text{-mole}^{-1} \, {}^{o}K^{-1}$)

T = temperature (${}^{o}K$)

An analysis of variance indicated no lack of fit when the linearized form of the model was applied either to the separate or combined sludge system rate data over a range of 5° to $25^{o}C$ at each individual system solids retention time of four, seven and ten days. The Arrhenius models (Table 4) for the separate and combined sludge data are illustrated in Figures 9, 10 and 11, together with the models for the nitrification reactor (B2) of the separate sludge system (Plant B). A system solids retention time of four, seven and ten days in the separate sludge system corresponds to values of six, ten and fifteen days in the nitrification reactor. The decrease in activation energy (E) with increased solids retention time indicates reduced temperature sensitivity at high SRT values (Table 4). The decrease may be caused by different relative changes in the nitrifying population present for a given temperature change for systems at different sludge ages. The importance of defining the system SRT in stating nitrification rates is illustrated by the variation in the Arrhenius parameter values (Table 4). The difference in rates observed by other authors (Figure 12) may be due to differences in SRT.
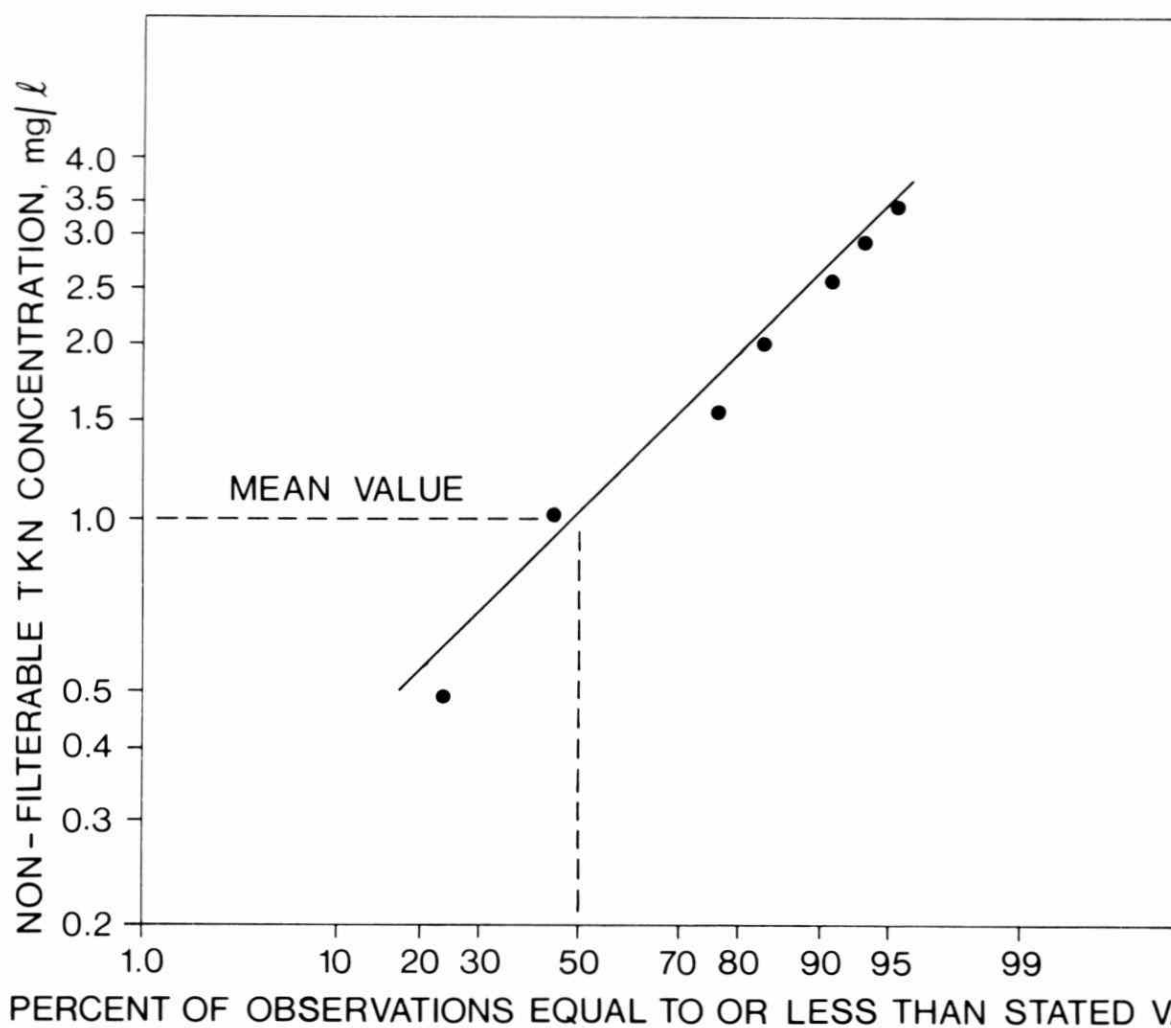
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FIGURE 8. PROBABILITY DISTRIBUTION FOR NON-FILTERABLE
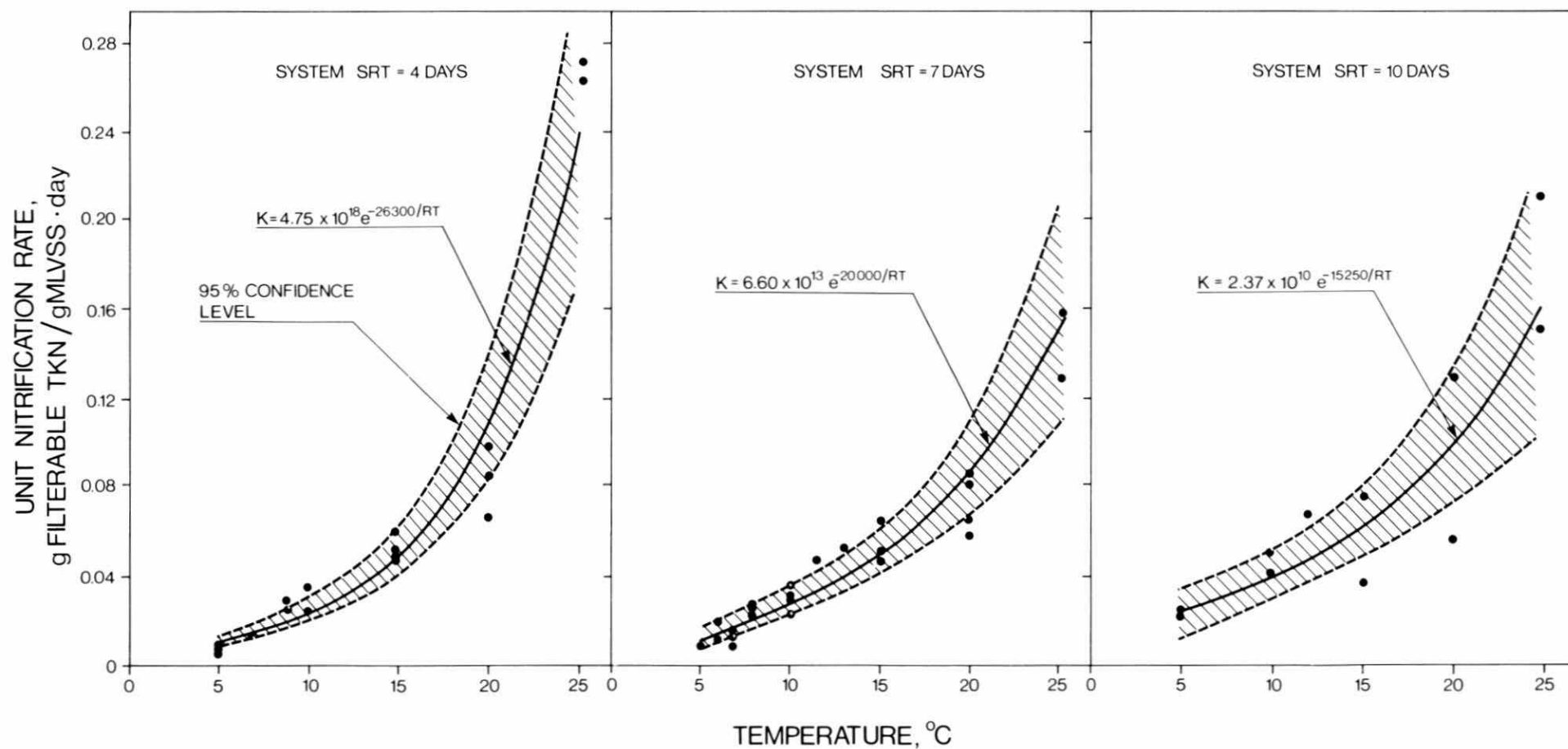EFFLUENT TKN FROM CARBON REMOVAL-NITRIFICATION
SYSTEMS

FIGURE 9. TEMPERATURE DEPENDENCY OF CARBON REMOVAL-NITRIFICATION IN SEPARATE SLUDGE SYSTEM
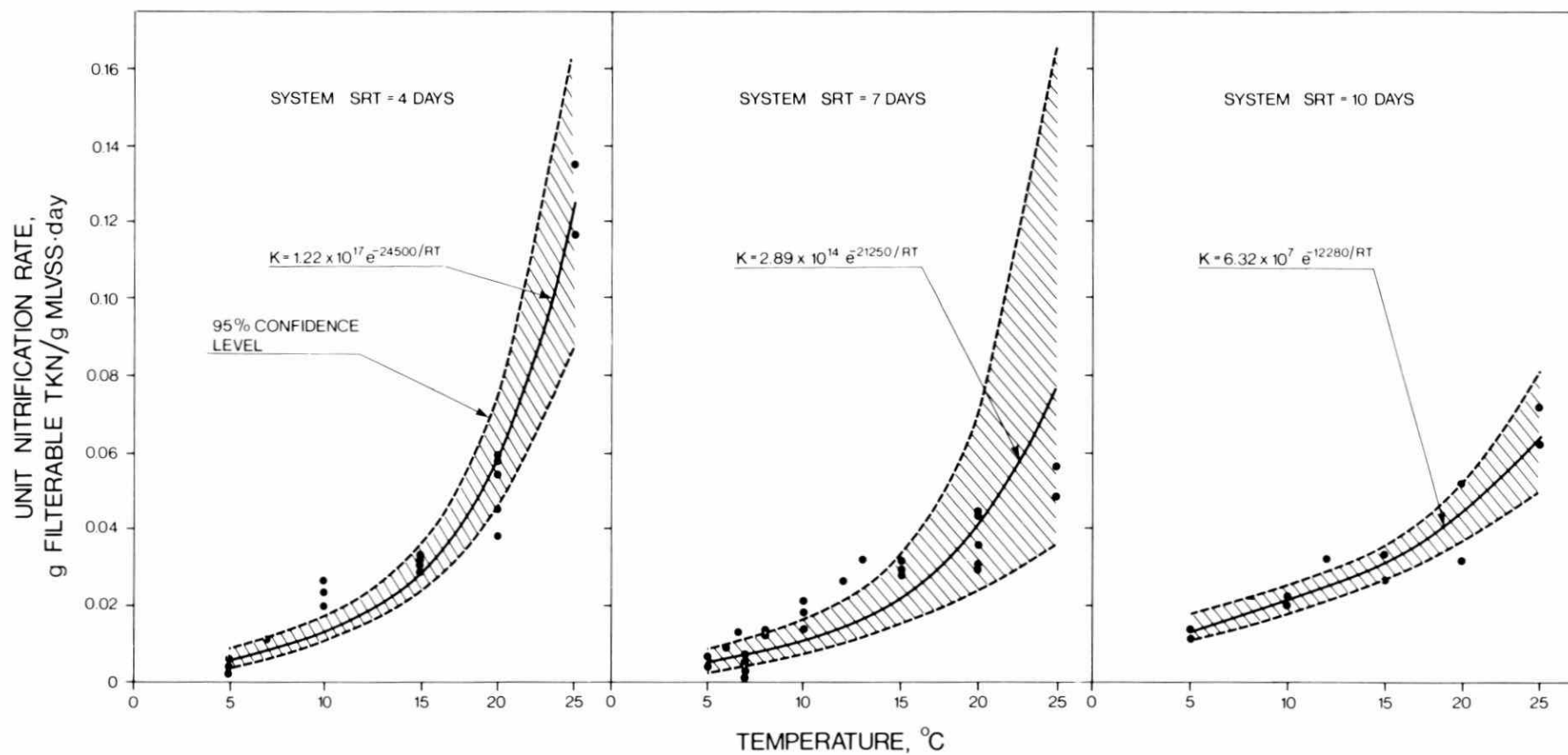
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FIGURE 10. TEMPERATURE DEPENDENCY OF CARBON REMOVAL-NITRIFICATION IN THE COMBINED SLUDGE SYSTEM
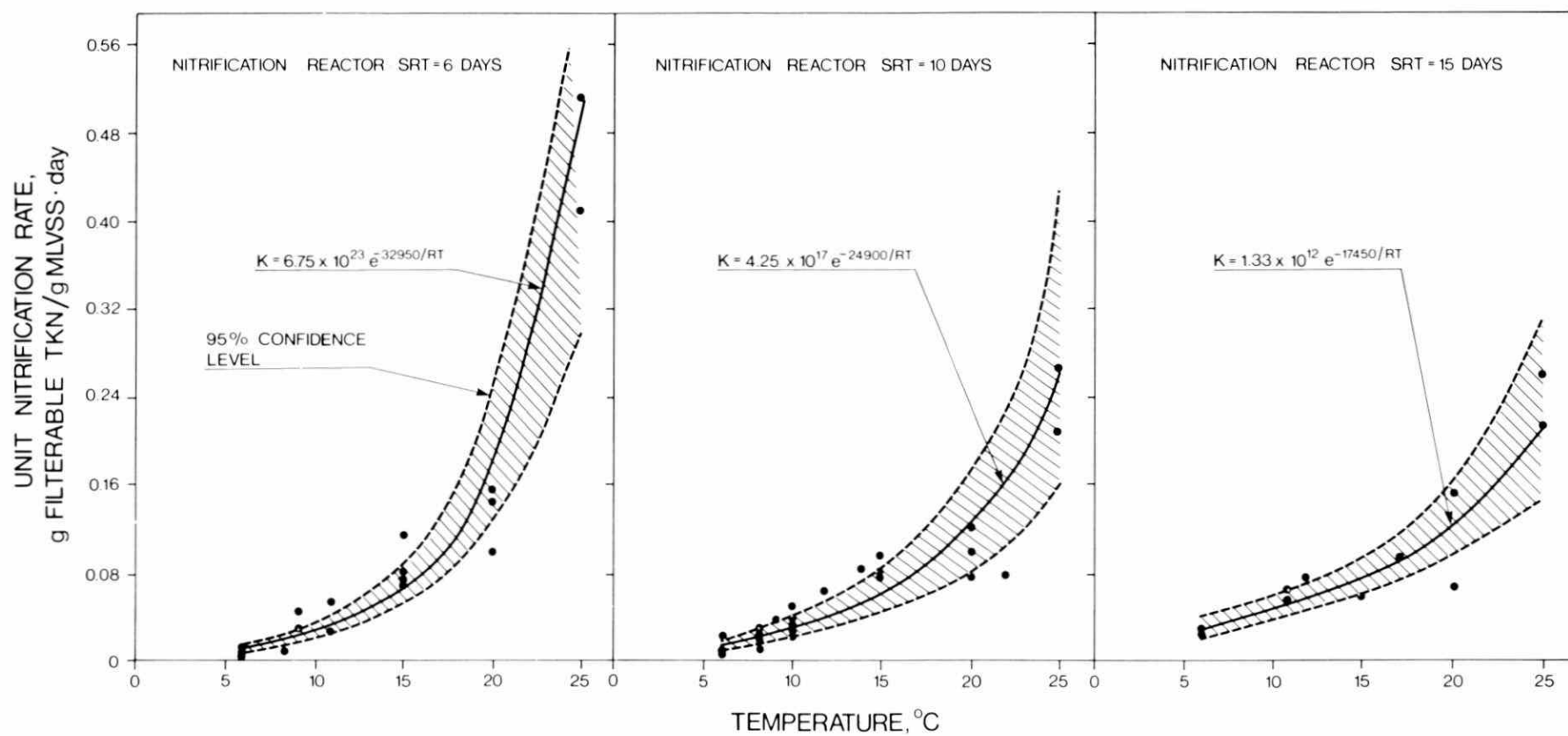
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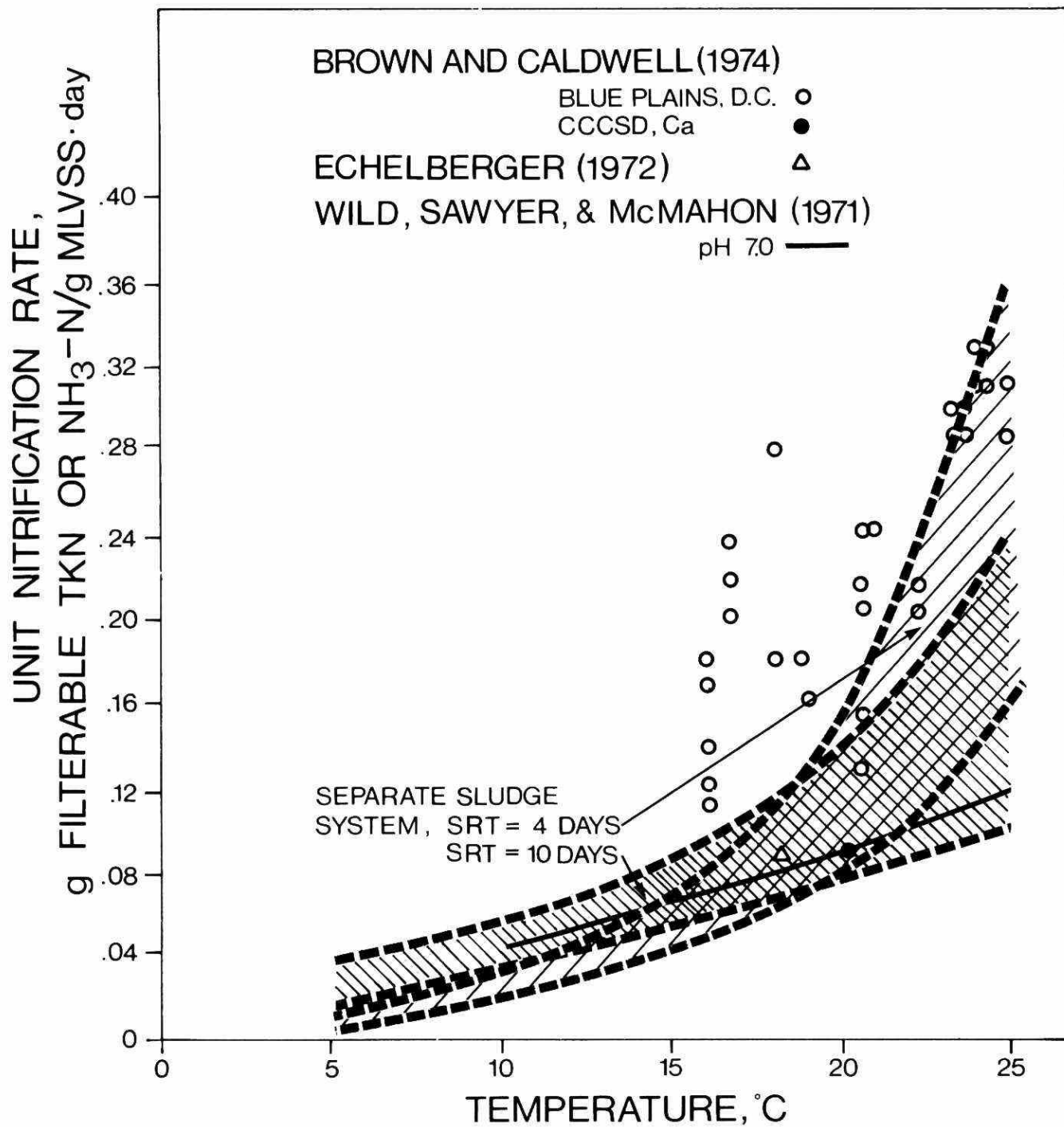FIGURE 11. TEMPERATURE DEPENDENCY OF SEPARATE STAGE NITRIFICATION

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FIGURE 12. NITRIFICATION RATES REPORTED BY VARIOUS AUTHORS

TABLE 4.   NITRIFICATION RATE VARIATION WITH TEMPERATURE IN SUSPENDED
GROWTH SYSTEMS

Reactor Configuration	Temp. Range °C	Arrhenius Model Parameters	
		E cal/g-mole	A
Carbon Removal-Nitrification System			
Separate Sludge			
4 day SRT	5-25	26,300	4.75×10^{18}
7 day SRT	5-25	20,000	6.60×10^{13}
10 day SRT	5-25	15,250	2.37×10^{10}
Combined Sludge			
4 day SRT	5-25	24,500	1.22×10^{17}
7 day SRT	5-25	21,250	2.89×10^{14}
10 day SRT	5-25	12,280	6.32×10^{7}
Nitrification System			
6 day SRT	6-25	32,950	6.75×10^{23}
10 day SRT	6-25	24,900	4.52×10^{17}
15 day SRT	6-25	17,450	1.33×10^{12}

Significant changes in raw wastewater pH are encountered
periodically at municipal treatment plants receiving quantities of
industrial wastes.   Variations in temperature may be anticipated in
treatment facilities in continental climates.   The reported sensitivity
of nitrification to temperature and pH (Wild, Sawyer, and McMahon,
1971) emphasizes the importance of defining the tolerance of such
systems to changes in these factors.   Parallel operation afforded a
direct qualitative comparison of the response and recovery profiles of
the separate sludge and single-stage combined sludge systems.   The
reactors were subjected to a step-down in temperature (15° to 10°C) and
a reduction in pH (HCl) by adjustment of the influent conditions.
Previous to each experiment, both were operated at solids retention
times of three to four days and at a temperature of approximately 15°C
(conditions critical for nitrification) to ensure measurable responses
would be observed.

A slower response to a step-down in temperature is indicated (Figure 13) for the separate sludge system. The approach by this system to TKN values predicted from pseudo "steady-state" results lagged behind the combined sludge system. The minimum attainable filterable TKN concentrations from the combined and separate sludge systems are 0.9 and 0.8 mg/l respectively. This corresponds to complete ammonia-nitrogen conversion. The change in effluent filterable COD concentration further indicated that for the separate sludge system there was a reduced response to changes in temperature.

The response to a reduction in feed pH was a significant change in effluent pH, and filterable TKN concentration (Figure 14). The separate sludge system exhibited a buffering capacity to a change in system pH which is reflected by the lag in response and recovery in terms of filterable effluent TKN concentration. This might be explained by the difference in system hydraulics. The nitrifying reactor (B2), of the separate sludge system exhibited a pH profile (Figure 14) having a greater magnitude and a shorter duration than that of the combined sludge reactor. The "predicted" increased effluent filterable TKN reflects the loss in reactor volatile solids caused by decreased settleability. The observed effluent filterable COD concentrations (Figure 14) indicate the reduced effect of the toxic conditions on the separate sludge system.

Any advantages of the separate sludge system such as a reduced response to changes in temperature must be balanced against the additional cost of added clarification facilities and increased solids production. The solids production from the parallel operating alternatives was assessed by determining the cumulative solids wasted during a pseudo "steady-state" operating period (Sutton, Murphy, Jank, and Monaghan, 1975). An average of 1.6 kg of solids was wasted from the separate sludge system for every 1 kg wasted from the combined system (Figure 15).
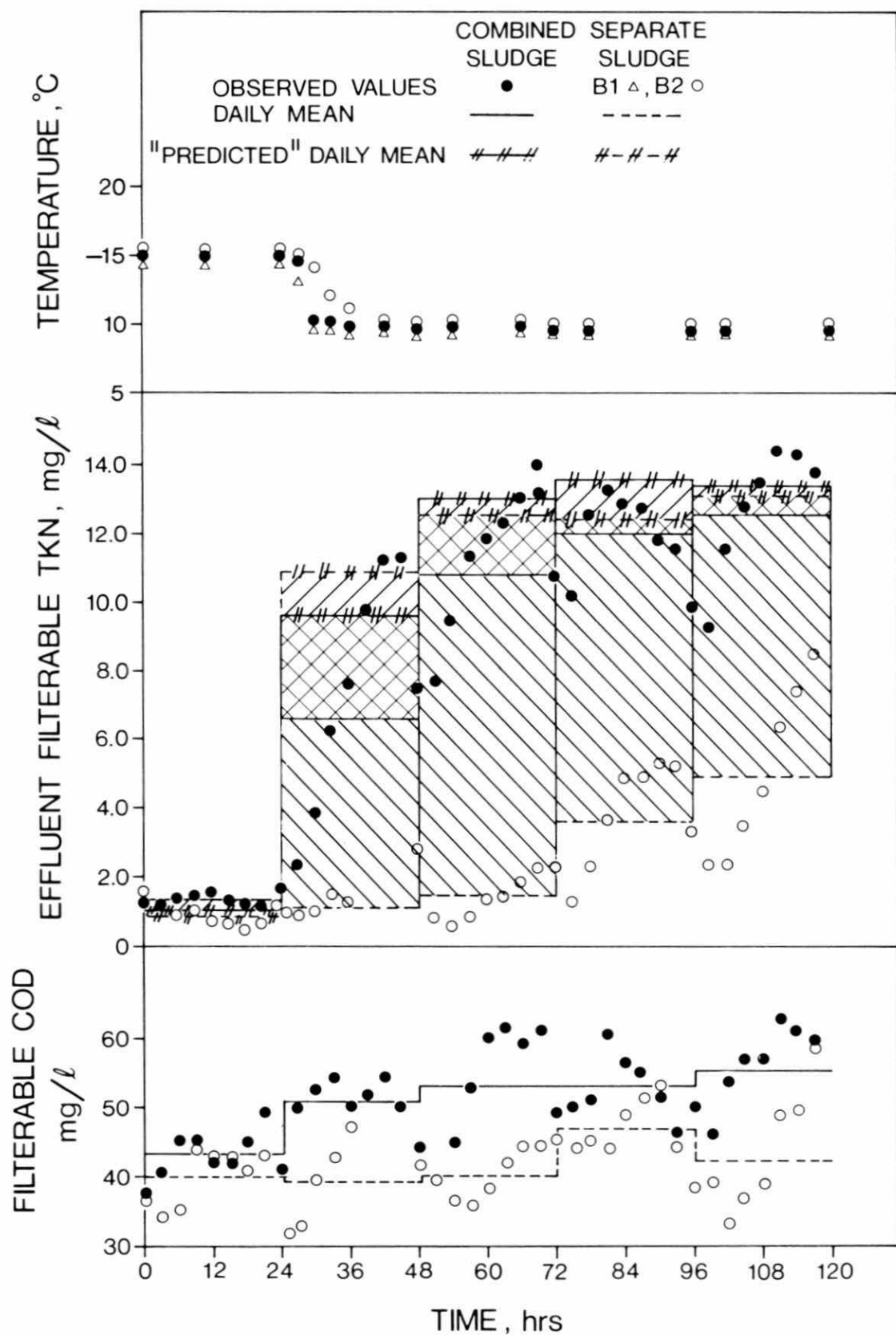
FIGURE 13. RESPONSE OF CARBON REMOVAL−NITRIFICATION
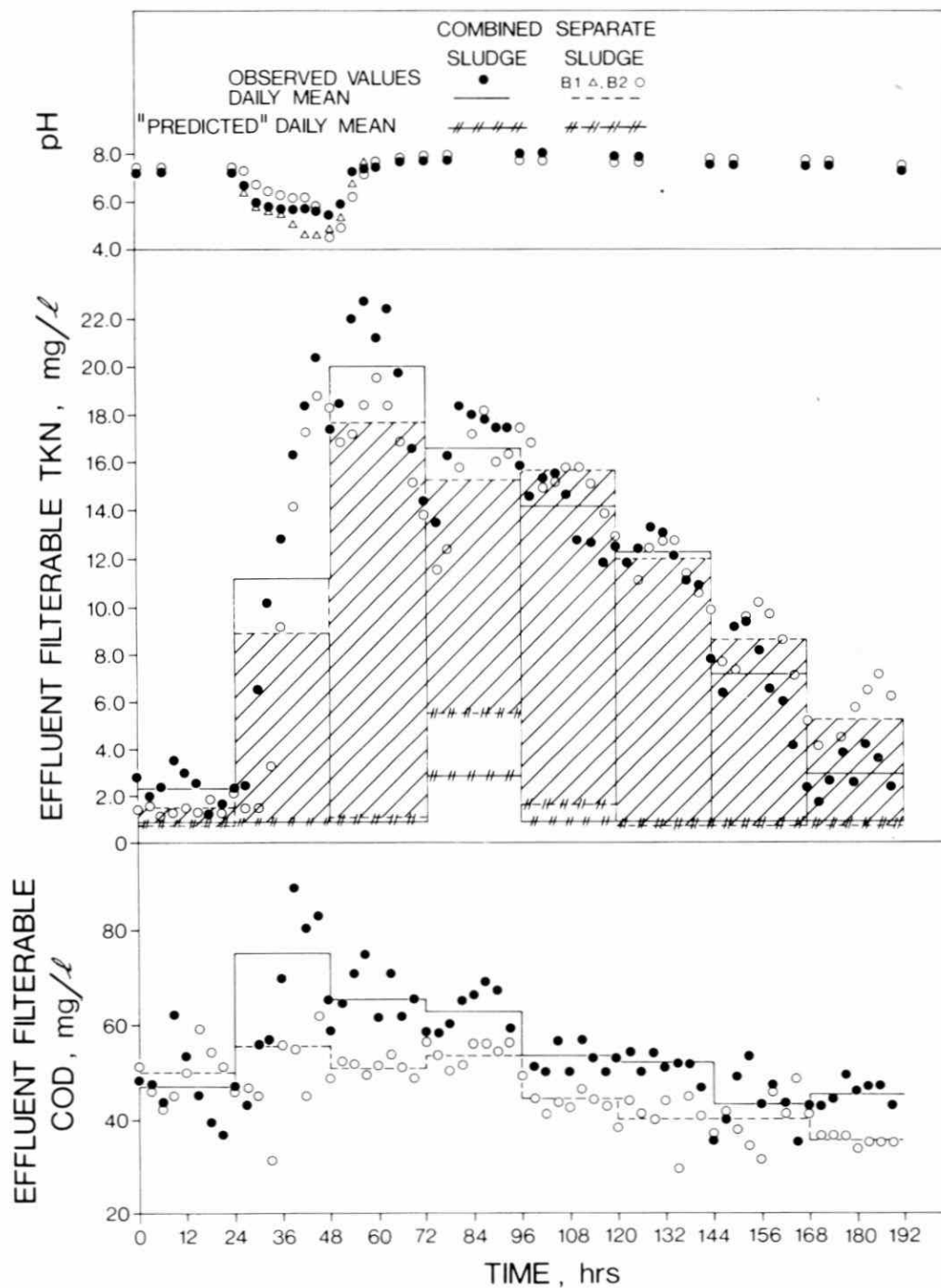SYSTEMS TO TEMPERATURE STEP−DOWN

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FIGURE 14. RESPONSE OF CARBON REMOVAL-NITRI-
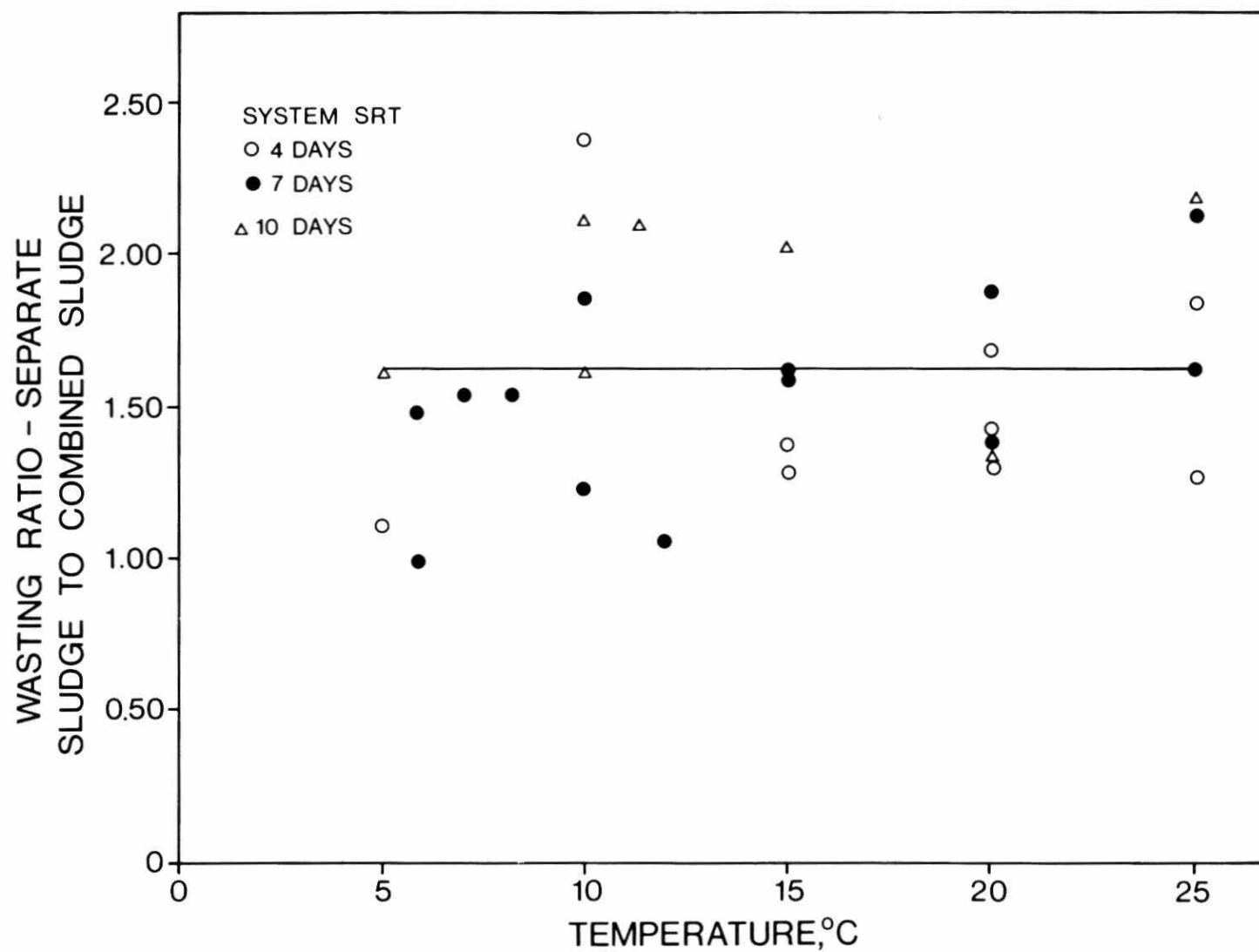FICATION SYSTEMS TO pH IMPULSE

FIGURE 15. SOLIDS WASTED IN PARALLEL SEPARATE AND COMBINED
CARBON REMOVAL-NITRIFICATION SLUDGE SYSTEMS

SUPPORTED GROWTH SYSTEM

Although the rotating biological contactor (RBC) has recently received attention for carbon removal and nitrification, little information exists to allow a comprehensive comparison of the system to suspended growth reactors. To define the specific capabilities of the system a significant amount of operational and kinetic data, over a temperature range of 7° to $25^{\circ}C$ has been gathered under conditions that afford direct and sensitive comparisons.

The zero-order nature of the nitrification reaction, with respect to NH_3-N concentration, allowed a simple reaction rate expression to be employed in the suspended growth systems. In supported growth systems the influence of diffusion, and therefore substrate concentration, would logically play an important role in determining reaction kinetics. Harremoes and Reimer (1975) proposed half-order kinetics to describe denitrification rates in a packed column. Haug and McCarty (1972) showed that the relationship between ammonia removal and substrate concentration in a submerged aerobic nitrification packed column was approximately first-order. Torpey, Heukelekian, Kaplovsky, and Epstein (1972) found a zero-order dependence in studying nitrification with a RBC.

To determine the removal rate expressions for half and first-order kinetics a hydraulic model is required. Although the pilot plant RBC was composed of four compartments in series, tracer analyses revealed that a model of two equal sized completely mixed tanks (CSTR's) more adequately described the actual flow conditions (Figure 16). Expressing the nitrification rate as filterable TKN removed per unit surface area of disc, attempts were made to fit zero, half, and first-order kinetic expressions. For the data obtained in this study the simple zero-order model adequately described the results (Wilson, Murphy, Sutton, and Jank, 1975). The lack of rate concentration dependency (zero-order) is clearly demonstrated for the $20^{\circ}C$ experimental conditions (Figure 17).

Operating the RBC over a temperature range of 7° to $25^{\circ}C$ and at a carbon loading between 6 and 12 g $BOD_5/m^2 \cdot day$ (1.22 and 2.45 lb
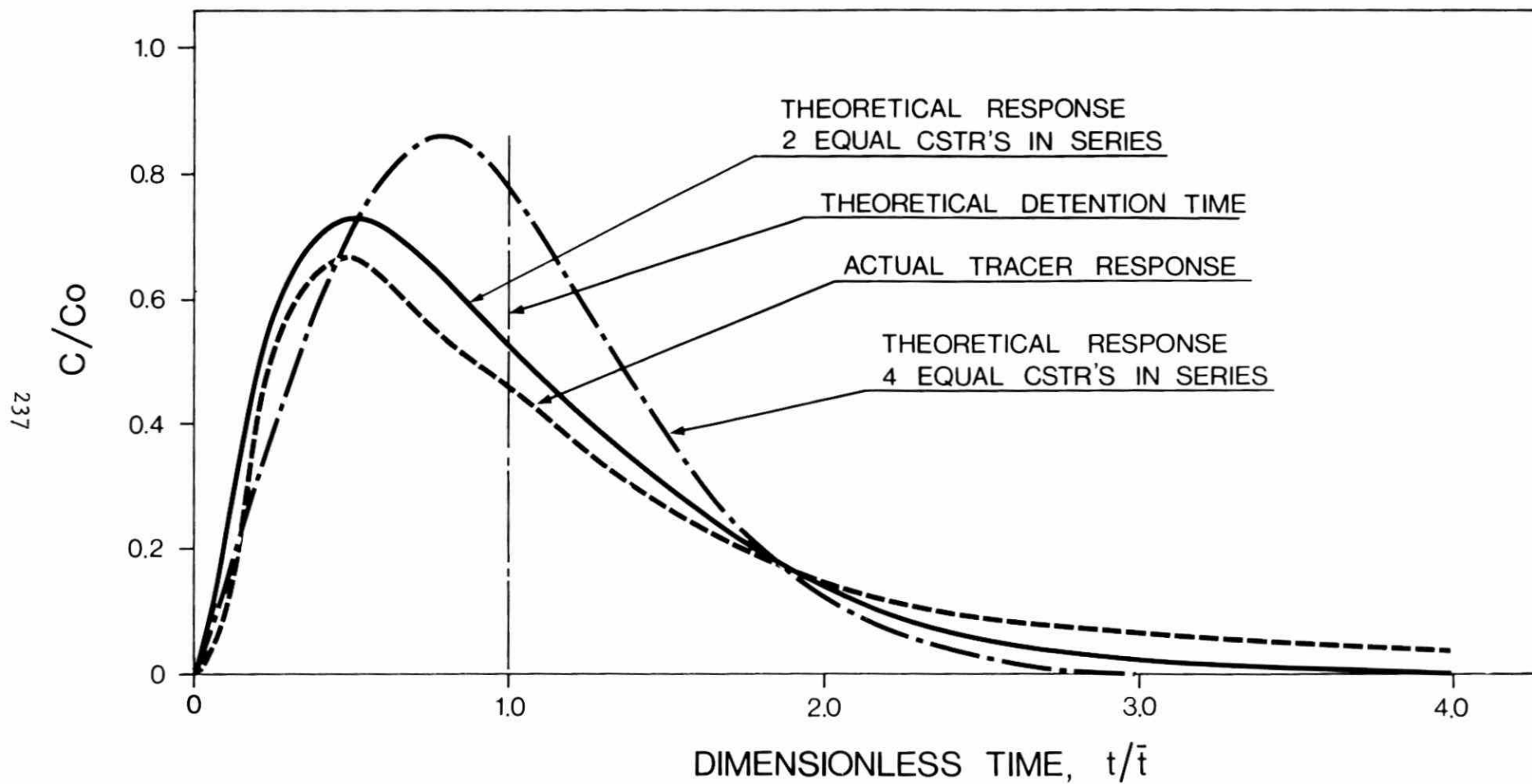
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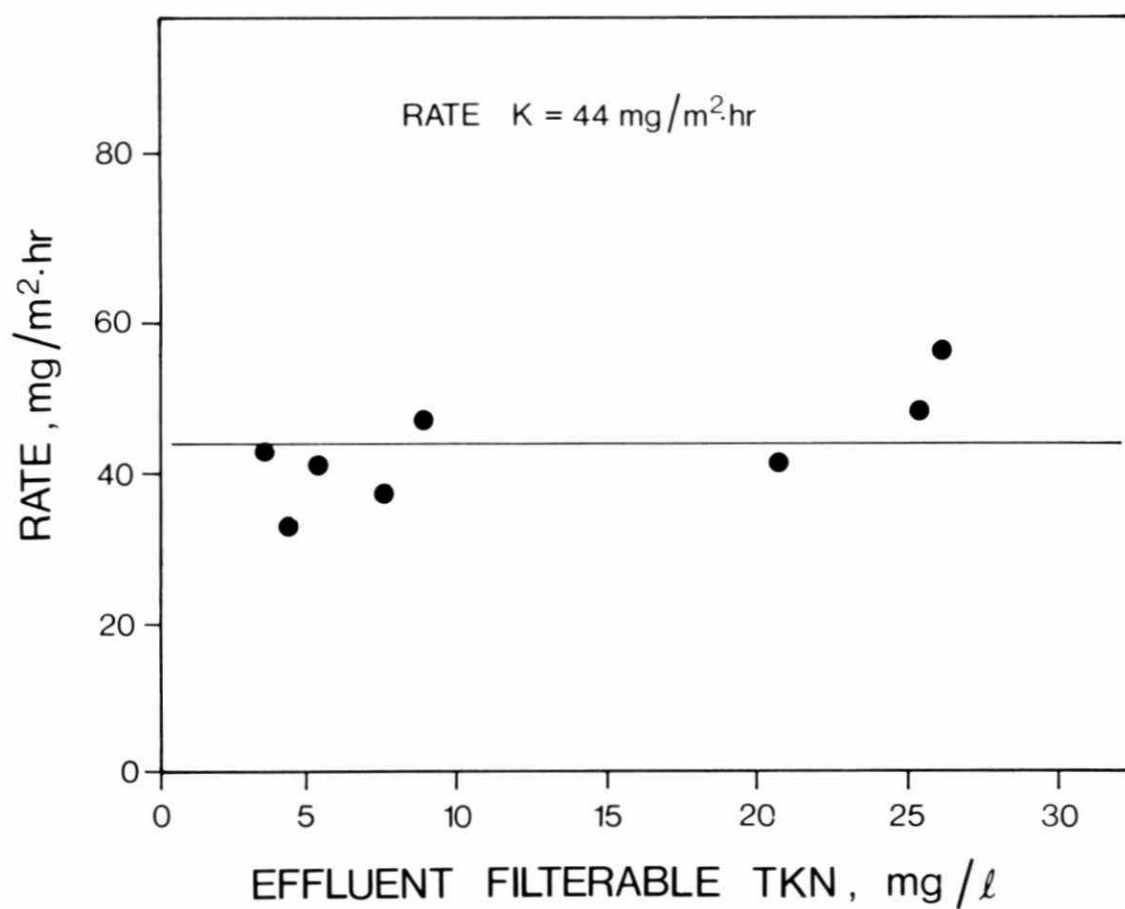
FIGURE 16. RBC FLOW CHARACTERISTICS BY TRACER ANALYSIS (PULSE INPUT)

FIGURE 17. NITRIFICATION IN THE RBC AT 20°C (ZERO-
ORDER KINETIC MODEL)

$BOD_5/1000$ ft$^2 \cdot$day), an Arrhenius model successfully described the temperature dependency of nitrification (Figure 18). The value of the activation energy (E) suggests a temperature sensitivity comparable to a suspended growth system with a SRT of 10 days (Figure 19).

A volumetric comparison provides an opportunity to compare the filterable TKN removal efficiency in the suspended and supported growth systems. Based on the volume of disc media an advantage is evident for the suspended growth system (Table 5). Another method of expressing the comparison is in terms of the volatile suspended solids (MLVSS) required in the suspended growth system in order to achieve a removal rate equivalent to that of the RBC (Table 5).

TABLE 5. VOLUMETRIC REMOVAL EFFICIENCY IN SUSPENDED AND SUPPORTED NITRIFICATION SYSTEMS AT 20°C

Nitrification System	Volumetric Removal g filterable TKN/m$^3 \cdot$day	Equivalent MLVSS mg/l
Combined Sludge SRT = 10 days (MLVSS = 2500 mg/l)	110	-
RBC*	92	2091

*Removal based on volume of disc media

Recognition of the heat transfer properties of the RBC necessitated the installation of a temperature control system for the air above the discs in order to maintain the experimental temperature level. This allowed heat losses to be measured for a variety of waste-water and ambient air conditions leading to development of the following model to describe the heat transfer process (Wilson, 1975):

$$Q = VC_p (T_{in} - T_{out}) = KAt (T_{ave} - T_{air}) \qquad (10)$$

where
Q = heat transfer in the RBC (kcal hr^{-1})

V = liquid volume of RBC (l)

C_p = specific heat of wastewater (kcal l^{-1} °C^{-1})
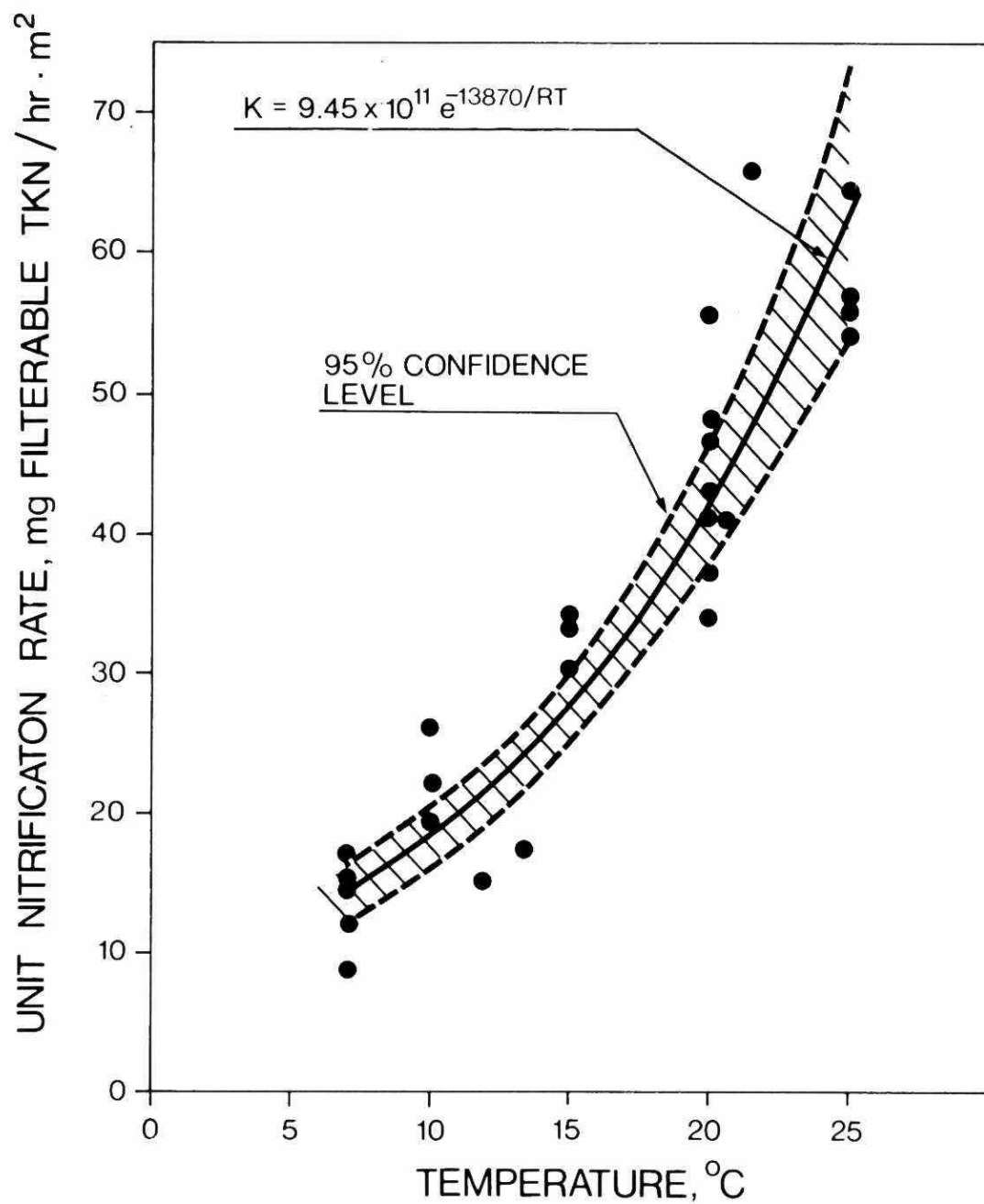
T_{in} = influent wastewater temperature (°C)

FIGURE 18. TEMPERATURE DEPENDENCY OF CARBON
REMOVAL-NITRIFICATION IN THE RBC

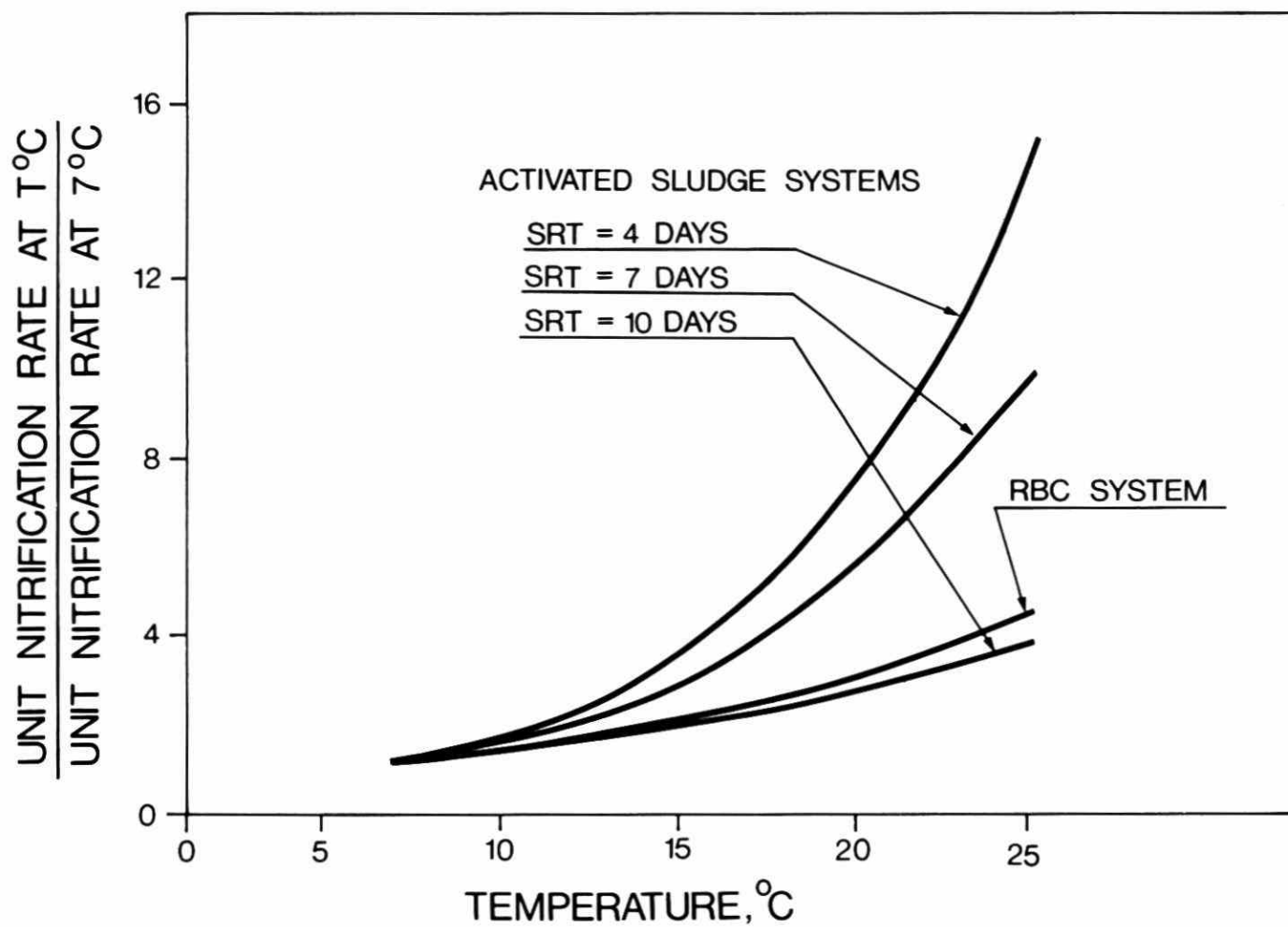FIGURE 19. CARBON REMOVAL–NITRIFICATION RATE DEPENDENCY
ON TEMPERATURE IN SUSPENDED AND SUPPORTED
GROWTH SYSTEMS

T_{out} = effluent wastewater temperature ($^{\circ}$C)

T_{ave} = (T_{in} + T_{out})/2

T_{air} = air temperature ($^{\circ}$C)

A = disc surface area (m^2)

t = hydraulic detention time of wastewater (hr)

K = heat transfer coefficient (kcal m^{-2} $^{\circ}C^{-1}$ hr^{-1})

This expression states that the rate of heat transfer between the wastewater in the RBC and the air above the discs is directly proportional to the total surface area of the discs and the mean temperature driving force between the wastewater and the air. Figure 20 is a graphical representation of equation 10 when the ambient air temperature is zero $^{\circ}$C. It shows the expected change in wastewater temperature for different conditions of influent temperature and hydraulic flow. The results indicate the necessity for providing a protective environment in areas where winter temperatures could lower the wastewater temperature sufficiently to cause severe decreases in biological activity.

DESIGN CONSIDERATIONS

Biological carbon removal-nitrification has been shown to be feasible employing suspended or supported growth systems under all temperature conditions normally encountered in municipal wastewater treatment. In designing nitrification systems the temperature, pH, alkalinity, and the presence of potential inhibitors will affect removal rates. Adequate dissolved oxygen (greater than 2.0 mg/l) and sufficient solids retention time must be maintained within the process reactor to ensure that nitrification will occur.

In designing for nitrification in suspended or supported growth systems specific considerations are necessary.

Suspended Growth Systems

1.   For nitrogen levels commonly found in domestic wastewater, the rate of nitrification, expressed as the filterable TKN removed per unit mass of activated sludge, is independent of the concentration of filterable TKN.
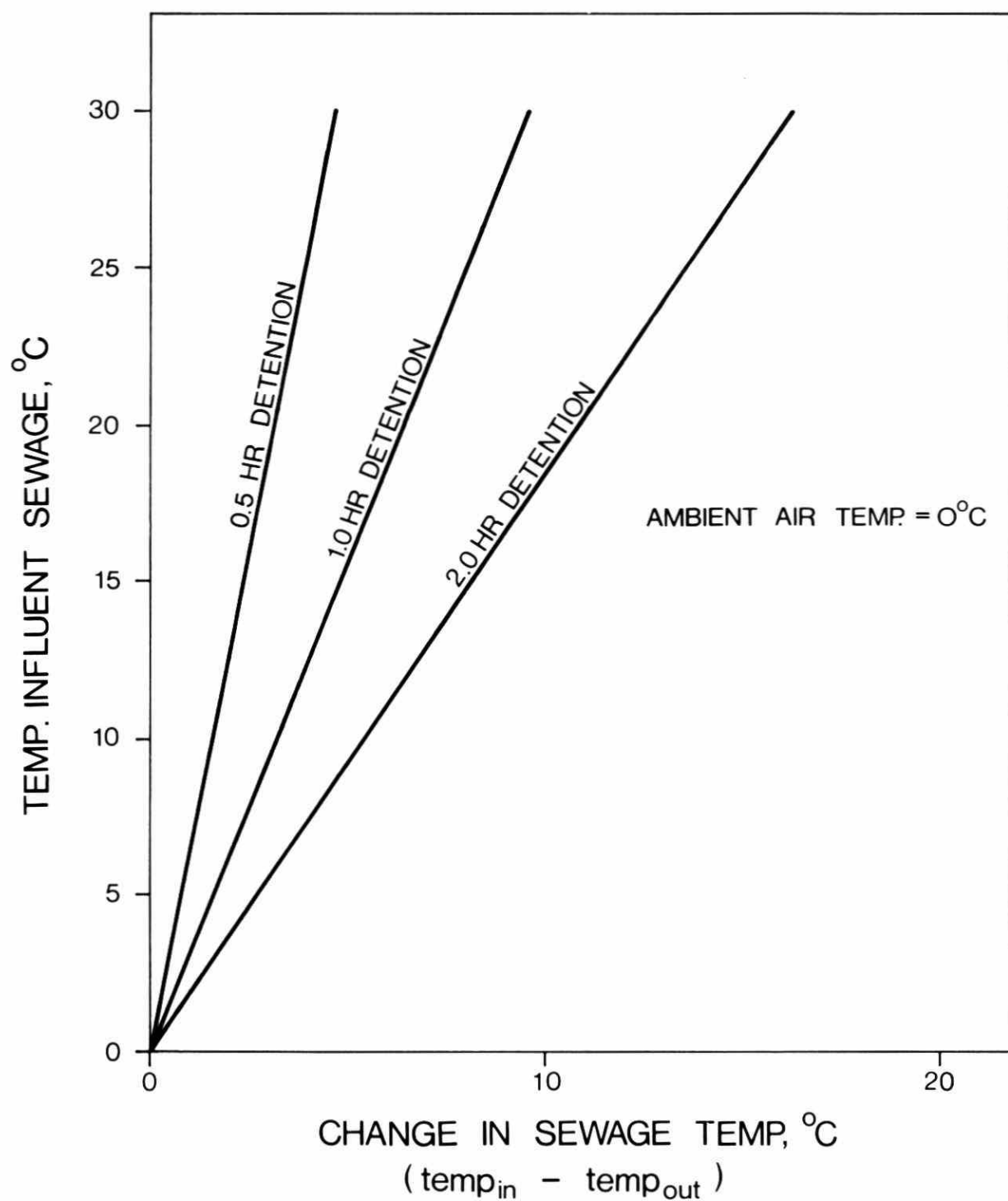
FIGURE 20. HEAT TRANSFER IN THE RBC

2.   No significant difference in filterable TKN removal or temperature sensitivity may be expected between systems with separate nitrifying sludges compared to combined carbon removal-nitrification sludges under similar operating conditions (i.e. temperature, solids retention time, etc.).

3.   Nitrification can be accomplished over a range of solids retention times with temperature sensitivity decreasing with increasing SRT.

4.   No appreciable difference in settleability was evident between the combined and separate nitrifying sludge systems. Even with complete nitrification, slightly less than 1 mg/l of filterable organic nitrogen together with an average of 1 mg/l of TKN may be expected in the suspended fraction of the effluent.

5.   For a step-down in temperature, the increase in effluent filterable TKN concentration for a separate sludge system will be considerably slower than that of a combined sludge system.

6.   A pulse change in influent pH will cause a lower reactor pH, but of shorter duration, in a separate sludge system. The accompanying increase in effluent concentration of both filterable TKN and COD will be less.

7.   A significantly higher quantity of sludge will be produced from a separate sludge carbon removal-nitrification system than from a combined sludge system.

Supported Growth System

1.   The rate of nitrification in the RBC is independent of concentration of filterable TKN and can be expressed as a function of the disc surface area within the limits of BOD_5 and filterable TKN loadings in this study.

2.   The temperature sensitivity of nitrification in the RBC is comparable to combined sludge carbon removal-nitrification systems at high SRT's.

3.   In cold climates wastewater cooling as a result of heat
          transfer between the RBC and the atmosphere will be significant.
          The resulting effect on efficiency will necessitate either
          the installation of the RBC's in a partially heated building,
          or the placement of covers on all units.

ACKNOWLEDGEMENTS

REFERENCES

Anderson, J.H., "Studies on the Oxidation of Ammonia by Nitrosomonas",
Biochem. J., 95, 684 (1965).

Antonie, R.L., "Nitrification and Denitrification with the Bio-Surf
Process", Presented at the Annual Meeting of the New England Wat. Pollut.
Control Association, Kennebunkport, Maine, (1974).

Brandon, T.H. and Grindley, J., "Effects of Nitrates on the Rising of
Sludge in Sedimentation Tanks", Journal Proc. Inst. of Sewage Purification,
175 (1944).

Brown and Caldwell, "Case Histories of Nitrification and Denitrification
Facilities" EPA Technology Transfer Design Seminar for Wastewater
Treatment Facilities, Orlando, Florida (1974).

Brown and Caldwell, "Process Design Manual for Nitrogen Control", U.S. EPA,
Office of Technology Transfer, Washington, D.C. (1975).

Busch, A.W., Aerobic Biological Treatment of Waste Waters, Principles and Practice, Oligodynamics Press, Houston, Texas, Chapt. 3, p. 94 (1971).

Buswell, A.M., Shiota, T., Lawrence, N. and Meter, I.V., "Laboratory Studies on the Kinetics of the Growth of Nitrosomonas with Relation to the Nitrification Phase of the BOD Test", Appl. Micro., 2, 21 (1954).

Deppe, K. and Engel, H., "Untersuchungen über Temperaturabhängigkeit der Nitratbildung durch Nitrobacter winogradski Buch. bei ungehemmten und gehemmten Wachstum", Zentbl. Bakt. Parasitkde II, 113, 561 (1960).

Downing, A.L., Painter, H.A. and Knowles, G., "Nitrification in the Activated Sludge Process", J. Inst. Sew. Purif., 130 (1964).

Downing, A.L., Tomlinson, T.G. and Truesdale, G.A., "Effect of Inhibitors on Nitrification in the Activated Sludge Process", Institute of Sewage Purification Journal and Proceedings, 537 (1964).

Engel, M.S., "The Growth and Aktotrophic Metabolism of Nitrosomonas europaea", Ph.D. Thesis, Cornell University, Ithaca, N.Y. (1959).

Harremoes, P. and Reimer, M., "Pilot Scale Experiments on Downflow Filter Denitrification", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Haug, R.T. and McCarty, P.L., "Nitrification with Submerged Filters", J. Wat. Pollut. Control Fed., 44, 11, 2086 (1972).

Huang, C.S., "Kinetics and Process Factors of Nitrification on a Biological Film Reactor", Ph.D. Thesis, State University of New York, Buffalo, N.Y. (1973).

Huang, C.S. and Hopson, N.E., "Nitrification Rate in Biological Processes", Jour. Envir. Eng. Div., Proceedings of the American Soc. of Civil Engineers, 100, 409 (1974).

IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Jeffrey, E.A. and Morgan, P.F., "Oxygen Demand of Digested Sludge Liquor", Sew. Ind. Wastes, 31, 1, 20 (1959).

Kiff, R.J., "The Ecology of Nitrification/Denitrification in Activated Sludge", Water Pollution Control, 71, 475 (1972).

Knowles, G., Downing, A.L. and Barrett, M.J., "Determination of Kinetic Constants for Nitrifying Bacteria in Mixed Culture with the Aid of an Electronic Computer", J. gen. Microbiol., 38, 263 (1965).

Laudelout, H. and van Tichelen, L., "Kinetics of the Nitrite Oxidation by Nitrobacter winogradski", J. Bact., 79, 39 (1960).

Lawrence, A.M. and Brown, C.G., "Biokinetic Approach to Optimal Design and Control of Nitrifying Activated Sludge Systems", Presented at the Annual Meeting of the New York Wat. Pollut. Control Association, New York, (1973).

Painter, H.A., "Microbial Transformations of Inorganic Nitrogen", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Stenquist, R.J., Parker, D.S. and Dosh, T.J., "Carbon Oxidation-Nitrification in Synthetic Media Trickling Filters", J. Wat. Pollut. Control Fed., 46, 10, 2327 (1974).

Sawyer, C.N. and Bradney, L., "Rising of Activated Sludge in Final Settling Tanks", Sewage Works Journal, 17, 6, 1191 (1945).

Sutton, P.M., "Suspended Growth Carbon Removal-Nitrification Systems", Ph.D. Thesis, McMaster University, Hamilton, Canada (1975) (in press).

Sutton, P.M., Murphy, K.L. and Jank, B.E., "Nitrogen Control: A Basis for Design with Activated Sludge Systems", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Sutton, P.M., Murphy, K.L., Jank, B.E. and Monaghan, B.A., "Biological Nitrogen Removal: The Efficacy of the Nitrification Step", J. Wat. Pollut. Control Fed., (1975) (in press).

Sutton, P.M. and Jank, B.E., "Design Considerations for Biological Nitrification - Denitrification Systems", EPS Technology Transfer Seminar on Alternatives for Nutrient Control, Kelowna, B.C. (1975).

Tenney, M.W. and Echelberger, W.F., "Removal of Organic and Eutrophying Pollutants by Chemical-Biological Treatment", EPA Report No. R2-72-076 (1972).

Torpey, W., Heukelekian, H., Kaplovsky, A.J. and Epstein, R., "Effects of Exposing Slimes on Rotating Discs to Atmospheres Enriched with Oxygen", Advances in Water Pollution Research, Pergamon Press, London (1972).

Wild, H.E., Sawyer, C.N. and McMahon, T.C., "Factors Affecting Nitrification Kinetics", J. Wat. Pollut. Control Fed., 43, 9, 1845 (1971).

Wilson, R.W., "Continuous Fixed Film Biological Nitrification and Denitrification of Wastewater", M.Sc. Thesis, McMaster University, Hamilton, Canada (1975).

Wilson, R.W., Murphy, K.L., Sutton, P.M. and Jank, B.E., "Nitrogen Control: Design Considerations for Supported Growth Systems", Presented at 48th Annual Conference of Wat. Pollut. Control Fed., Miami Beach (1975).

Wuhrmann, K. and Mechsner, K., "Beitrag zur Kenntnis der Mikrobiellen Denitrifikation", _Path. Microbiol._, _26_, 579 (1965).

Wuhrmann, K., "Effects of Oxygen Tension on Biochemical Reactions in Sewage Purification Plants", _Advances in Biological Waste Treatment_, Proceedings 3rd Conf. Biol. Waste Treatment, edited by W.W. Eckenfelder and J. McCabe, Pergamon Press, New York, p. 27 (1963).

DESIGN CONSIDERATIONS FOR BIOLOGICAL
DENITRIFICATION SYSTEMS

by

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ENVIRONMENT CANADA

INTRODUCTION

Several potential environmental problems resulting from the discharge of ammonia can be eliminated by upgrading existing biological treatment facilities to promote nitrification. Deleterious effects associated with the resulting nitrate-nitrogen may require installation of facilities for denitrification. Infant cyanosis is a serious problem resulting from high concentrations of nitrate in drinking water. Nitrification alone does not eliminate eutrophication problems associated with high nitrogen wastewater effluents. Industrial water reuse may be affected by high nitrate levels.

Biological denitrification is achieved under anaerobic conditions by heterotrophic microorganisms that utilize nitrate as a hydrogen acceptor when an organic energy source is available. A wide variety of common facultative bacteria such as Pseudomonas, Micrococcus, Achromobacter, Spirillum and Bacillus have been reported to accomplish denitrification (Delwiche, 1956).

BIOLOGICAL DENITRIFICATION

Principles

The reduction of nitrate occurs through a series of complicated enzyme catalyzed reactions which can follow either the assimilatory or dissimilatory route. In assimilatory denitrification, nitrate is reduced to ammonia that is used in cell synthesis. In
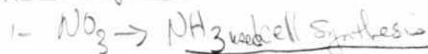
dissimilatory denitrification, in the absence of molecular O_2, nitrate serves as the hydrogen acceptor in the oxidation of the carbon substrate, producing energy for cell growth. The gaseous end products of the reaction are principally dinitrogen (N_2).

In oxygen respiration the same series of reversible enzymatic reactions are involved in transferring electrons from organic substrates to molecular oxygen. Only a different terminal enzyme is required for nitrate respiration and this is formed by the bacteria in the absence of oxygen. Consequently nitrate dissimilation is inhibited by oxygen. The occurrence of denitrification under aerobic conditions, as reported by some researchers (Myers and Matsen, 1955; Schmidt and Kampf, 1962), is most likely due to a dissolved oxygen gradient across the bacterial floc.

In wastewater treatment, denitrification normally will follow the processes of carbon removal and nitrification. Therefore, the residual carbon will not be of sufficient concentration, nor in a readily usable form for use by the denitrifying population as an organic carbon and energy source. The carbon and energy source may be provided internally by by-passing part of the raw sewage to the de-nitrification reactor or by the endogenous respiration of the sludge, or externally by the addition of a carbon and energy source. The use of raw sewage has an obvious economic advantage but results in ammonia appearing in the effluent from the denitrification system. Denitrification utilizing the endogenous respiration of sludge is feasible but the reaction rates obtained are approximately 25 percent of the rates found with systems utilizing external carbon and energy sources (Sutton, Murphy, and Jank, 1975). Various organic materials such as sugar, acetic acid, and acetone have been found satisfactory as external electron donors but methanol appears to be the optimum choice (McCarty, Beck, and St. Amant, 1969). An attractive alternative to methanol may be carbon-rich industrial wastes such as starch, sugar, and brewery production waste effluents.

If denitrification is considered to be a two-step process with methanol as the carbon and energy source the following reactions can be written:
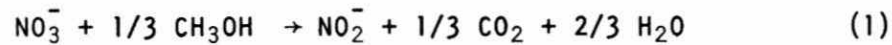
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Step 1:

$$NO_3^- + 1/3 \ CH_3OH \rightarrow NO_2^- + 1/3 \ CO_2 + 2/3 \ H_2O \qquad (1)$$

Step 2:

$$NO_2^- + 1/2 \ CH_3OH \rightarrow 1/2 \ N_2 + 1/2 \ CO_2 + 1/2 \ H_2O + OH^- \quad (2)$$

Overall:

$$NO_3^- + 5/6 \ CH_3OH \rightarrow 1/2 \ N_2 + 5/6 \ CO_2 + 7/6 \ H_2O + OH^- \quad (3)$$
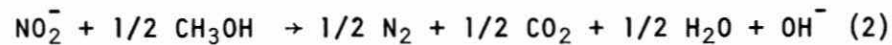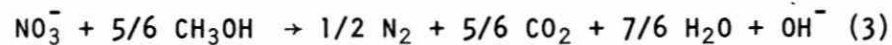
In addition to the methanol required for nitrate dissimilation (Equation 3), an amount is necessary for bacterial growth. An overall mass balance which considers this requirement is as follows (McCarty, 1973):

$$CH_3OH + 0.92 \ NO_3^- + 0.92 \ H^+ \rightarrow$$

$$0.06 \ C_5H_7O_2N + 0.43 \ N_2 + 0.7 \ CO_2 + 2.25 \ H_2O \qquad (4)$$
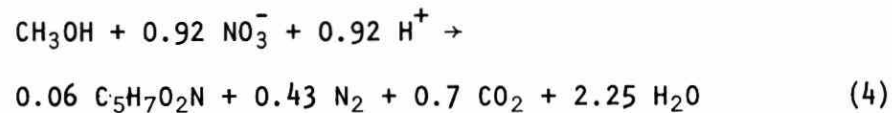
On the basis of this equation, 20 mg of NO_3-N would require 19 mg of CH_3OH-C and would produce 10.5 mg of bacteria. Denitrification studies have determined that the actual methanol requirement and biomass production can be estimated from the following relationship (McCarty, Beck, and St. Amant, 1969):

Methanol requirement:

$$C_m = 2.47 \ N_o + 1.53 \ N_1 + 0.87 \ D_o \qquad (5)$$

Biomass production:

$$C_b = 0.53 \ N_o + 0.32 \ N_1 + 0.19 \ D_o \qquad (6)$$

where: C_m =   required methanol (mg/l),

C_b =   biomass production (mg/l),

N_o =   initial NO_3-N concentration (mg/l),

N_1 =   initial NO_2-N concentration (mg/l), and

D_o =   the dissolved oxygen concentration in the wastewater entering the system (mg/l).

A number of researchers (Stensel, Loehr, and Lawrence, 1973; Sutton, Murphy, and Dawson, 1974) have obtained experimental results which

verify that these expressions provide reasonable estimates of methanol requirements and biomass production.

In denitrification, carbon and nitrate are the important substrates involved in the biological reaction. Numerous researchers have shown that the denitrification rate is independent of nitrate concentration (zero-order) down to values of less than 1.0 mg/l NO_3-N in suspended growth reactors (Moore and Schroeder, 1970; Murphy and Sutton, 1974). In supported growth systems both zero-order (Requa and Schroeder, 1973; Sutton, Murphy, and Dawson, 1975) and half-order (Harremoes, 1975) nitrate concentration dependencies have been reported. No denitrification rate dependence on carbon concentration will be observed if an external source is supplied in excess of the amount required to meet the needs of cell synthesis and nitrate reduction. If the desire is to maximize the reaction rate and hence minimize reactor size, then the denitrification process must be operated under conditions where carbon concentration is not limiting.

Environmental Factors

Reduced denitrification rates in the presence of dissolved oxygen can be accounted for by the effect of oxygen on the formation of necessary enzymes, or activity of preformed enzymes, required for dissimilatory denitrification (Chang and Morris, 1962). Although denitrification will be observed in systems at low oxygen tension, a completely separate anaerobic reactor will maximize nitrate reduction (Painter, 1975).

The optimum pH for denitrification is usually reported at conditions near neutrality (Dawson and Murphy, 1973). Nitrate dissimilation is a basic reaction (Equation 4) with hydrogen ion consumption tending to raise the pH. When used together with nitrification, the nitrification-denitrification sequence tends to maintain a neutral pH.

Denitrification is reported to occur over a temperature range from $0^{\circ}C$ to $50^{\circ}C$ with an optimum approaching $40^{\circ}C$ (Christensen and Harremoes, 1975). There is good agreement between authors

concerning the degree of temperature sensitivity over the range normally encountered in municipal wastewater treatment (Figure 1). An extensive list of values for various coefficients normally used to express the reaction rate and temperature dependency in biological systems is presented elsewhere (Sutton, Murphy, and Dawson, 1974).

In biological nitrification-denitrification the autotrophic nitrifiers are more sensitive to environmental conditions than the heterotrophic denitrifiers. Ensuring that there are no toxicants in the wastewater limiting the nitrification rate, will also ensure that the denitrification process will be unaffected.

Denitrification Treatment Alternatives

Denitrification can be achieved in suspended and supported growth reactors.

In suspended growth reactors denitrification can be obtained with a separate denitrifying sludge or using the same sludge which provided carbon removal and/or nitrification. Wuhrmann (1964) and Christensen, Harremoes, and Jensen (1975) have obtained high nitrogen removals in a single sludge system (Figure 2) using an internal carbon source. Mulbarger (1971) investigated combined sludge nitrification-denitrification preceded by separate sludge carbon or BOD removal (Figure 2).

A unique combined sludge system involving carbon removal-denitrification followed by nitrification (Figure 3) was first studied by Ludzack and Ettinger (1962). In order to promote high nitrogen removal in this system a large degree of mixed liquor recycle must be maintained.

Recently the oxidation ditch has received attention as a combined sludge nitrogen removal process (Matsché and Spatzierer, 1975). The aeration rotors (Figure 4) induce cycling of the mixed liquor creating aerobic and anaerobic zones and consequently promoting nitrification and denitrification. Another suspended growth system proposed for carbon removal-nitrification and denitrification is the alternating denitrification-nitrification process (Barnard, 1974). According to the author, the scheme consists of a series of anaerobic
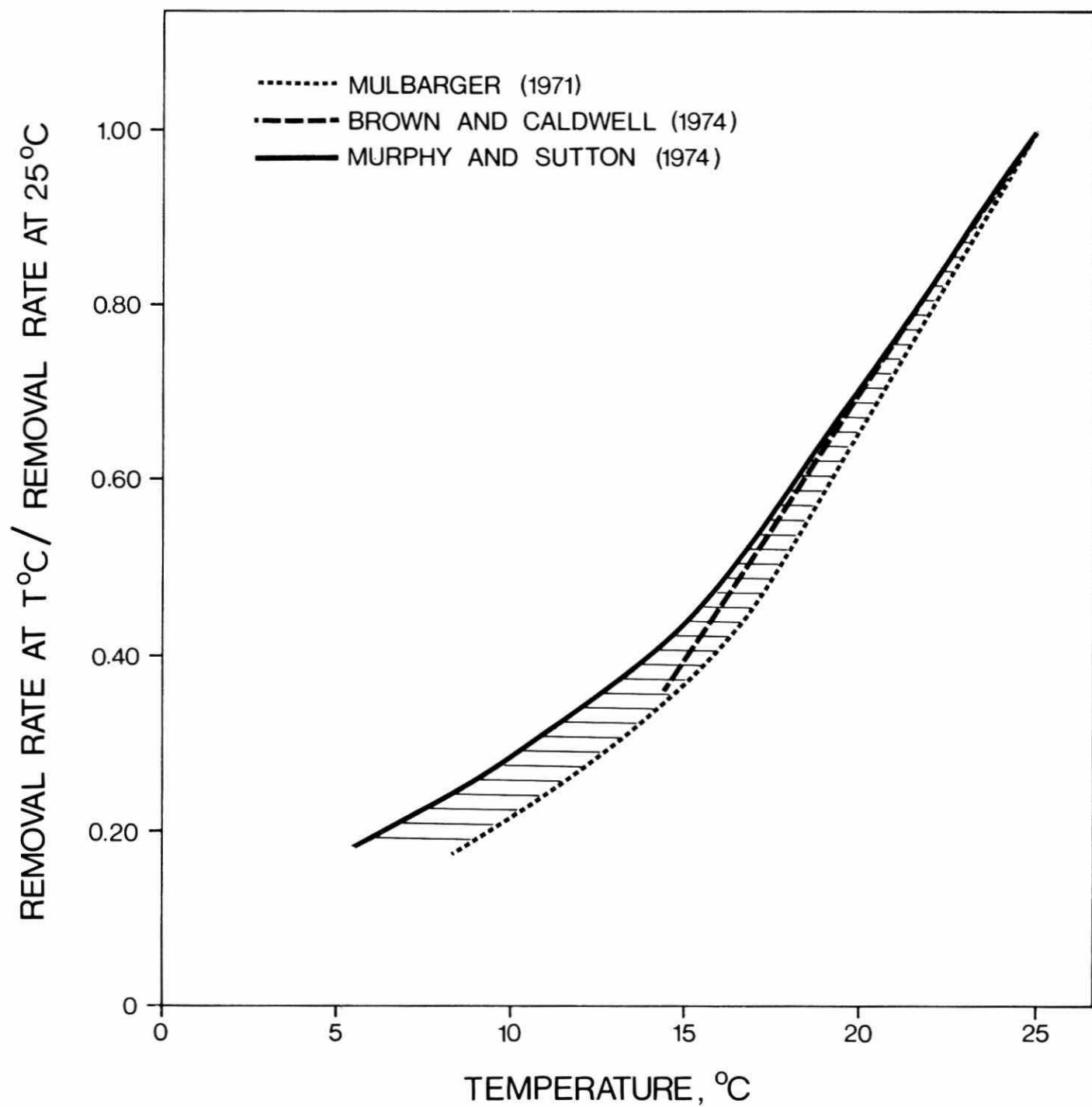
FIGURE 1. EFFECT OF TEMPERATURE ON DENITRIFICATION RATE

SINGLE SLUDGE
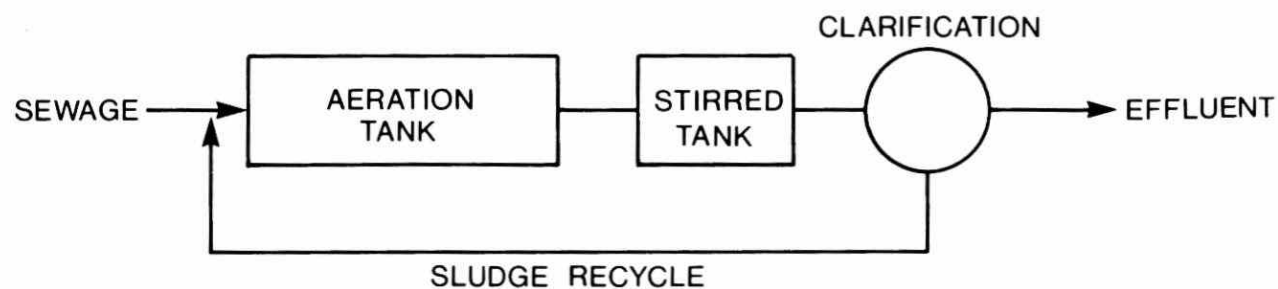
AERATION TANK: BOD REMOVAL AND NITRIFICATION
STIRRED TANK: NITRATE REMOVAL

BOD REMOVAL AND NITRIFICATION – DENITRIFICATION COMBINED SLUDGE
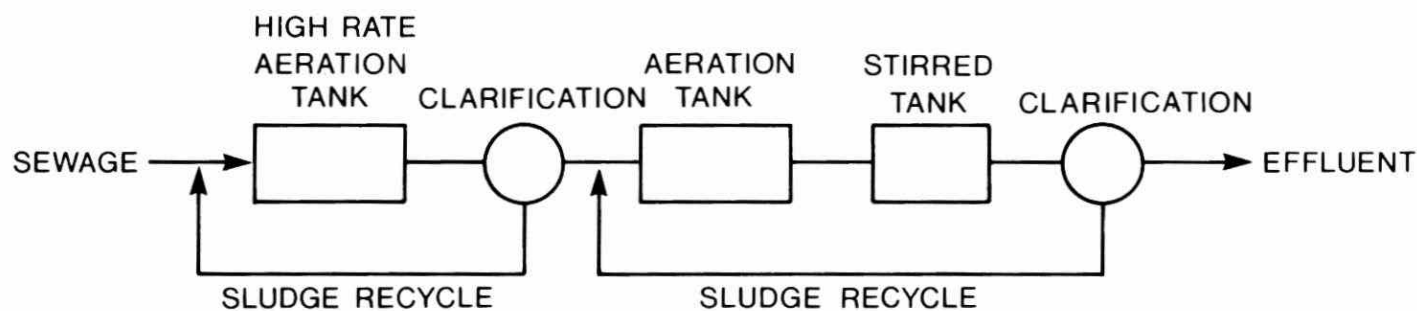
HIGH RATE AERATION TANK: BOD REMOVAL
AERATION TANK: NITRIFICATION
STIRRED TANK: NITRATE REMOVAL

FIGURE 2. COMBINED SLUDGE NITRIFICATION – DENITRIFICATION SYSTEMS

MIXED LIQUOR RECYCLE

SEWAGE → | STIRRED TANK | → | AERATION TANK | CLAR-IFICA-TION | → EFFLUENT

SLUDGE RECYCLE
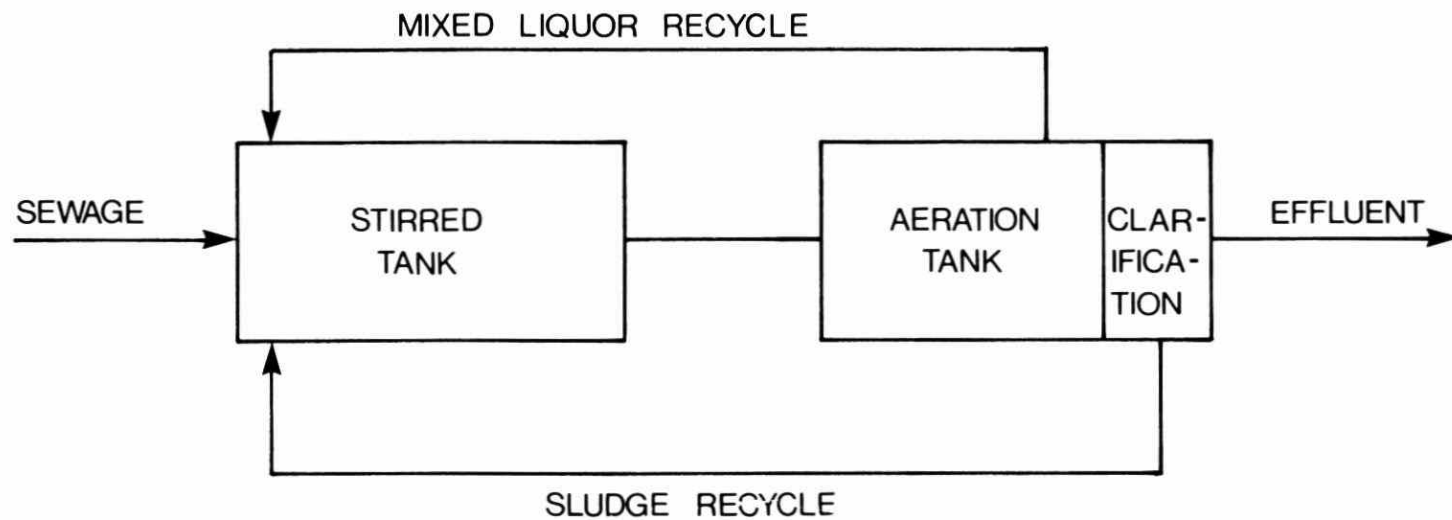
STIRRED TANK : NITRATE REMOVAL AND BOD REDUCTION
AERATION TANK : AMMONIA CONVERSION TO NITRATE AND
BOD REMOVAL

FIGURE 3. COMBINED SLUDGE CARBON REMOVAL – DENITRIFICATION AND
NITRIFICATION SYSTEM

OXIDATION DITCH

SEWAGE

SLUDGE RECYCLE

AERATION ROTORS

EFFLUENT

CLARIFICATION

AERATION TANKS:
BOD AND NITROGEN REMOVAL

ALTERNATING DENITRIFICATION - NITRIFICATION PROCESS

MIXED LIQUOR RECYCLE

CLARIFICATION

SEWAGE

STIRRED TANK

AERATION TANK A

STIRRED TANK

AERATION TANK B

EFFLUENT

SLUDGE RECYCLE

STIRRED TANKS: NITRATE AND BOD REMOVAL
AERATION TANK A: NITRIFICATION AND BOD REMOVAL
AERATION TANK B: NITRIFICATION AND SLUDGE STABILIZATION

FIGURE 4. PROCESS MODIFICATIONS FOR BIOLOGICAL NITROGEN REMOVAL

and aerobic basins (Figure 4) in which both nitrogen and phosphorus removal occur without chemical addition.

Together with other proposed combined sludge systems utilizing internal carbon sources, the aforementioned process schemes have not shown consistent, high nitrogen removal. This, together with reported low reaction rates (Barnard, 1974; Christensen, Harremoes, and Jensen, 1975) and corresponding long reactor detention times, may limit their application.

To date, separate sludge denitrification with the addition of an external carbon source has proven to be the most reliable nitrate removal system (Brown and Caldwell, 1975). High reaction rates and consistent performance have been reported by numerous authors (Mulbarger, 1971; Horstkotte, Niles, Parker, and Caldwell, 1974; Murphy and Sutton, 1974).

An optimal suspended growth system for nitrogen removal results when the separate sludge denitrification reactor is preceded by a separate (Figure 5) or combined sludge carbon removal-nitrification system.

A number of supported growth reactors consisting of columns containing different types of packed media are available for biological denitrification (Figure 6). The column systems generally fall into one of the following categories:

1) fixed bed upflow reactors with porous media of varying size exhibiting minor pressure gradients,

2) expanded bed upflow reactors exhibiting a constant pressure gradient, and

3) fixed bed downflow or upflow reactors exhibiting an increasing pressure gradient necessitating backwashing or flushing.

The column reactors of the first category have shown significant but variable nitrate removal (Wilson, 1975; Sutton, Murphy, and Dawson, 1974). Although only a limited quantity of design information is available, the expanded bed and pressure filters appear to be acceptable supported growth systems for denitrification (Jeris, Beer, and Mueller, 1974; English, Carry, Masse, Pitkin, and Dryden, 1974).
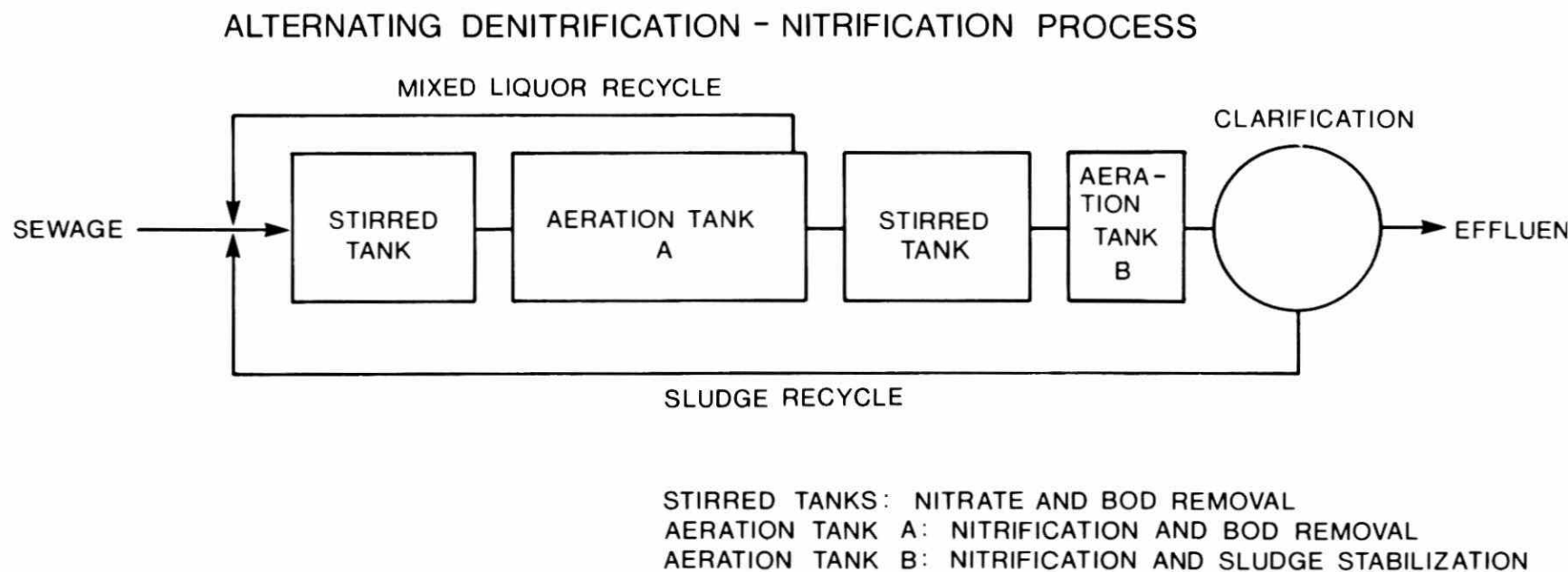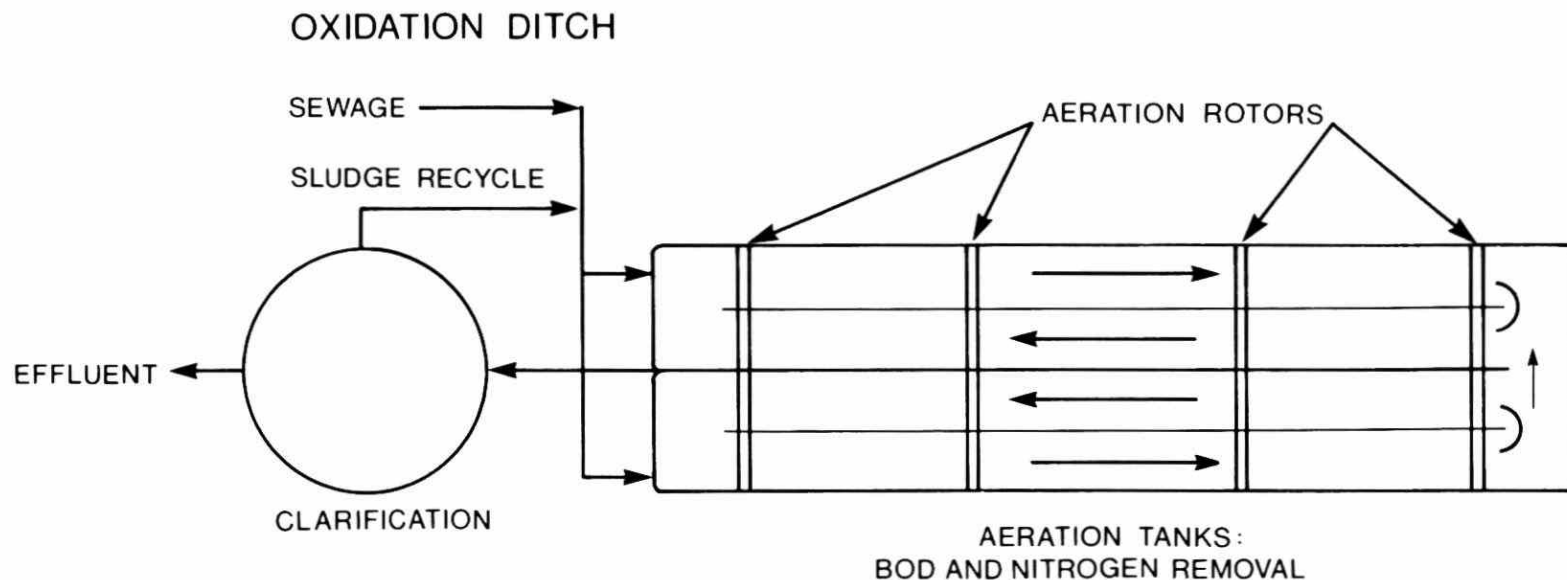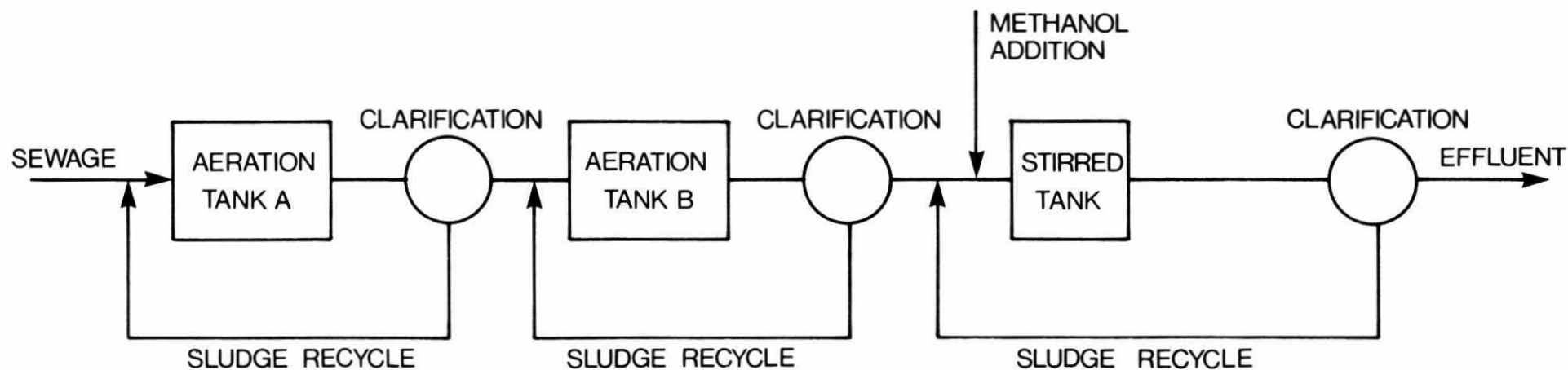
AERATION TANK A : BOD REMOVAL
AERATION TANK B : RESIDUAL BOD REMOVAL AND NITRIFICATION
STIRRED TANK : NITRATE REMOVAL

FIGURE 5. THREE-STAGE CARBON REMOVAL, NITRIFICATION AND DENITRIFICATION

SUBMERGED ANAEROBIC FILTER

EFFLUENT

PACKED TOWER

PACKED TOWERS: DENITRIFICATION

NITRIFIED INFLUENT

FLUSH WATER

CARBON SOURCE

DOWNFLOW FILTER

NITRIFIED INFLUENT

CARBON SOURCE

PACKED TOWER

EFFLUENT

BACKWASH WATER

EXPANDED BED CONTACTOR

SUBMERGED ROTATING BIOLOGICAL CONTACTOR

NITRIFIED INFLUENT

CARBON SOURCE

ROTATING DISCS

EFFLUENT

CLARIFICATION

WASTE SLUDGE

SOLIDS SEPARATION

EFFLUENT

SOLIDS RECYCLE

CARBON SOURCE

EXPANDED MEDIA

NITRIFIED INFLUENT

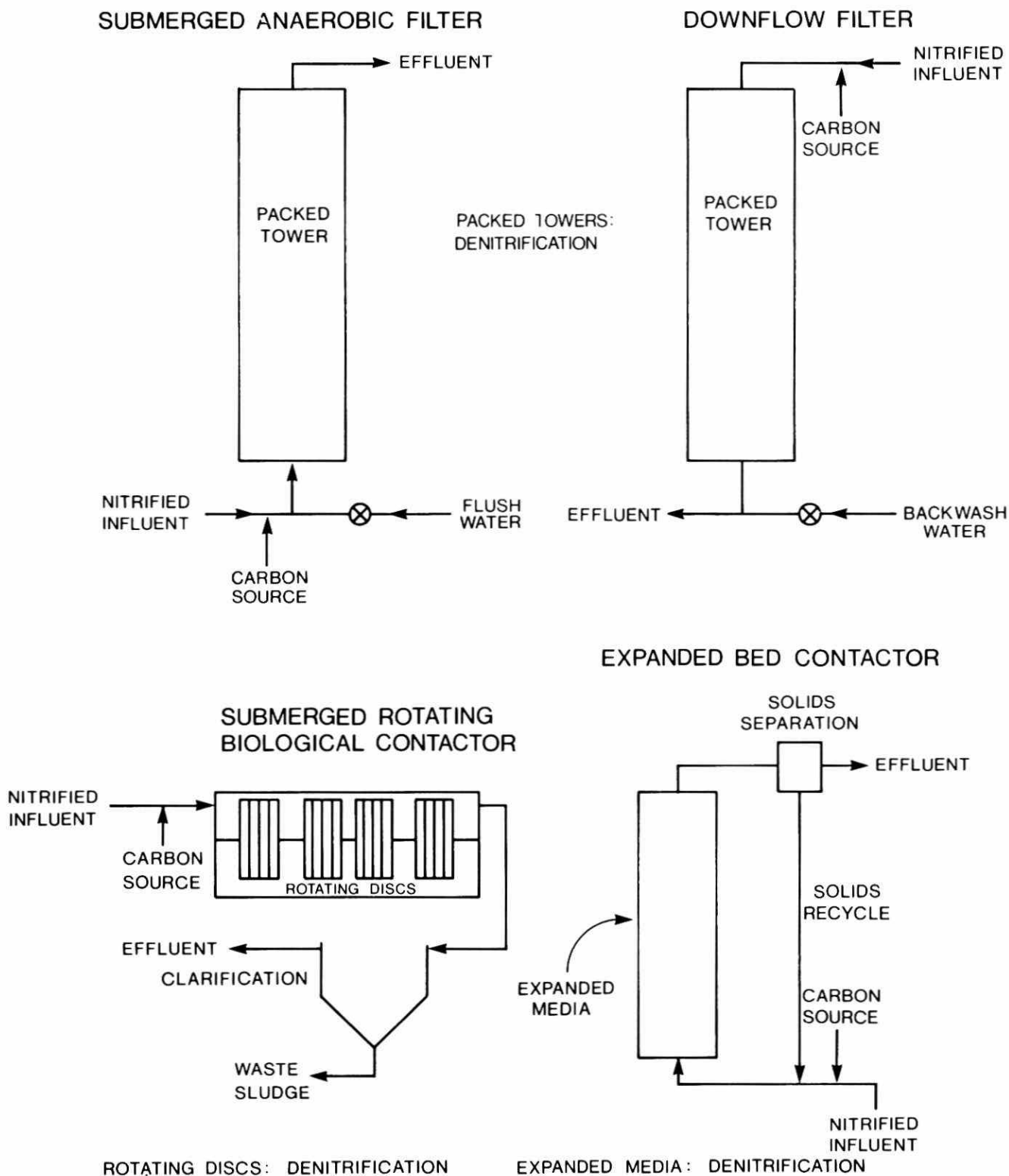ROTATING DISCS: DENITRIFICATION   EXPANDED MEDIA: DENITRIFICATION

FIGURE 6. SUPPORTED GROWTH DENITRIFICATION SYSTEMS

The rotating biological contactor can be employed for denitrification (Figure 6) by submerging the rotating discs below the liquid level to minimize oxygen transfer and maintain anaerobic conditions within the biological film. Preliminary results indicate significant nitrate removals are possible (Wilson, Murphy, Sutton, and Jank, 1975).

DENITRIFICATION PILOT PLANT PROGRAM

In May 1972, an intensive biological nitrification-denitrification research program began at the Wastewater Technology Centre, Burlington, Ontario, Canada. The pilot plant program was a joint project between McMaster University and the Environmental Protection Service. Design information was obtained for suspended and supported growth systems through operation of a 19.6 m^3/day (4320 Igpd) denitrification pilot plant (Figure 7).

The suspended growth denitrification system consisted of an anaerobic mixing tank followed by a dispersed air aeration, sludge stabilization tank and clarifier.

The supported growth system, packed columns or upflow filters for denitrification, consisted of units containing either ceramic Intalox saddles or plastic Pall rings. The Intalox saddles investigated were 0.95 cm (0.375 in) and 1.27 cm (0.5 in) providing surface areas of 787 and 623 m^2/m^3 (240 and 190 ft^2/ft^3) respectively with porosities of 78 and 70 percent. The series of three Intalox packed columns (Figure 7) had a total empty bed detention time of 24 minutes at a feed rate of 2.3 l/min (0.5 Igpm). A total of six columns were available, thus allowing the two different Intalox saddle sizes to be investigated simultaneously. The Pall rings investigated were 2.54 cm and 5.08 cm providing surface areas of 207 and 102 m^2/m^3 (63 and 31 ft^2/ft^3) respectively and porosities of 90 and 92 percent. The two Pall ring columns (Figure 7) each had an empty bed detention time of approximately 76 minutes at the applied hydraulic loading of 2.3 l/min (0.5 Igpm). Further details concerning the reactor systems are presented elsewhere (Sutton and Jank, 1975).

SUSPENDED GROWTH SYSTEM

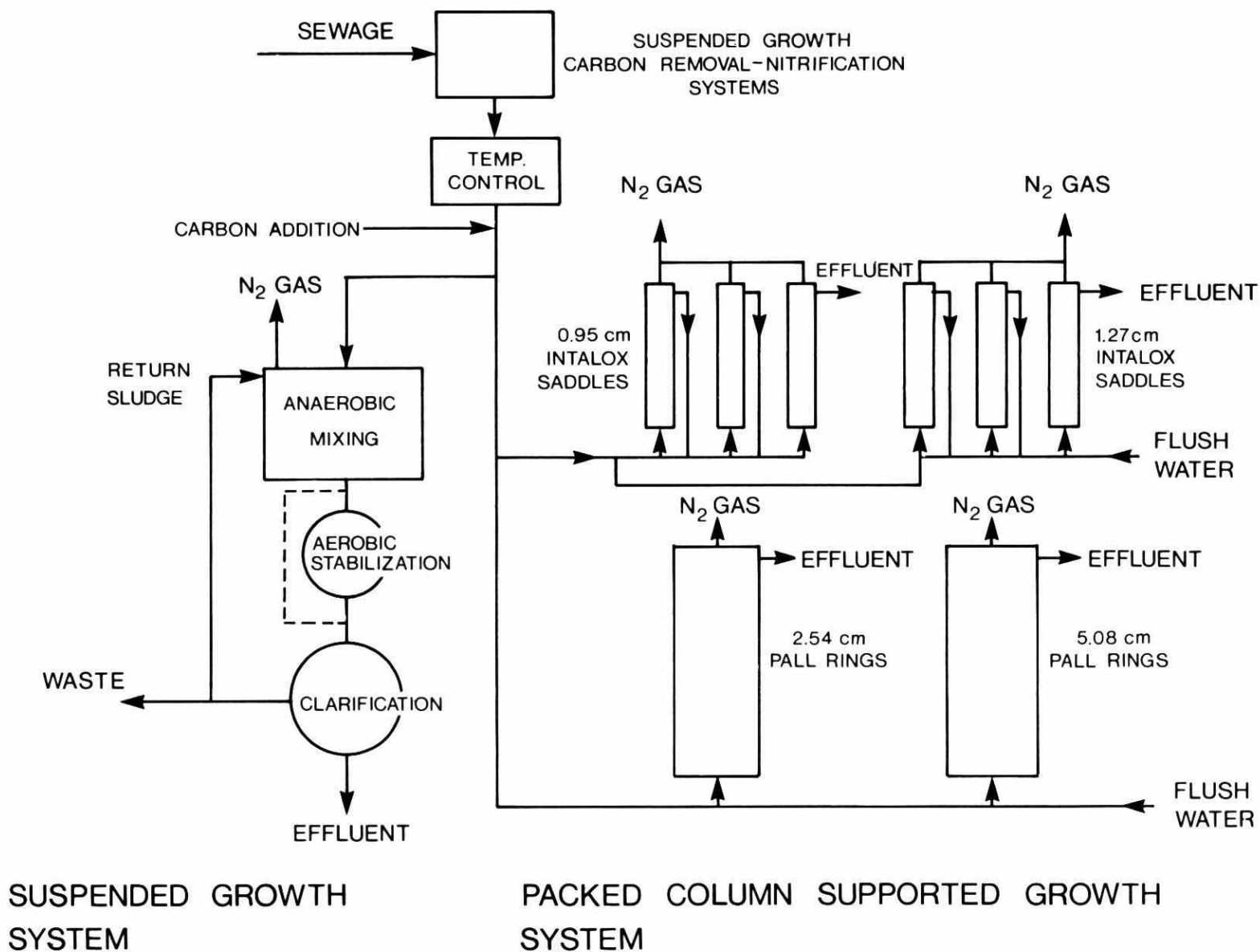PACKED COLUMN SUPPORTED GROWTH SYSTEM

FIGURE 7. SUSPENDED AND SUPPORTED GROWTH DENITRIFICATION PILOT PLANT

The denitrification pilot plant reactors received nitrified secondary effluent. The geometric mean values for nitrate plus nitrite-nitrogen and suspended solids concentrations were 10 and 34 mg/l respectively, (Figures 8 and 9).

In order to determine the kinetic removal rates used to describe the performance of the systems, the effluent from the reactors must contain residual nitrate-nitrogen. When complete nitrate removal was obtained, additional amounts of nitrate (KNO_3) were added. To avoid any carbon limitation on the nitrate removal rate, methanol was added to the influent of the denitrification reactors.

In addition to the above studies current investigations include studies of:

1) a supported growth rotating biological contactor for denitrification,

2) a supported growth downflow sand filter for denitrification, and,

3) an upflow supported growth expanded bed contactor for denitrification.

SUSPENDED GROWTH SYSTEMS

Denitrification rates were obtained during continuous operation of the separate sludge reactor at SRT's of three, six and nine days over a temperature range of 5° to 25°C. In addition, batch rate data was obtained at three and six day SRT levels.

The independence of the denitrification rate and nitrate plus nitrite concentration in suspended growth systems has been shown by numerous authors (Sutton, Murphy, and Dawson, 1974; Moore and Schroeder, 1970). The zero-order nature of the reaction allows use of a simple unit denitrification rate expression.

Describing the variation of rate with temperature by an Arrhenius expression, an analysis of variance indicated no lack of fit when this modelling procedure was applied either to the individual or combined batch and continuous rate data (Murphy and Sutton, 1974). As
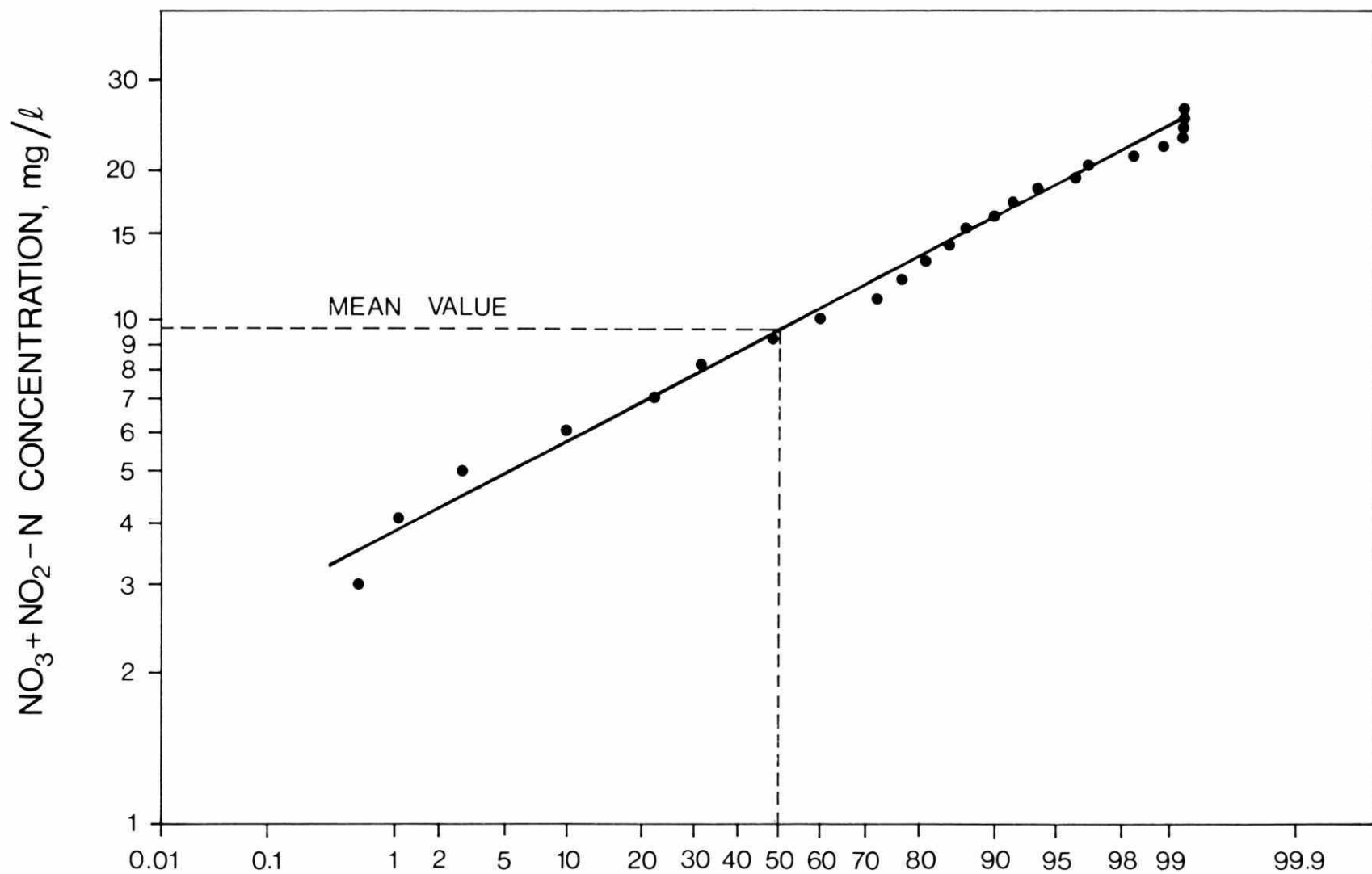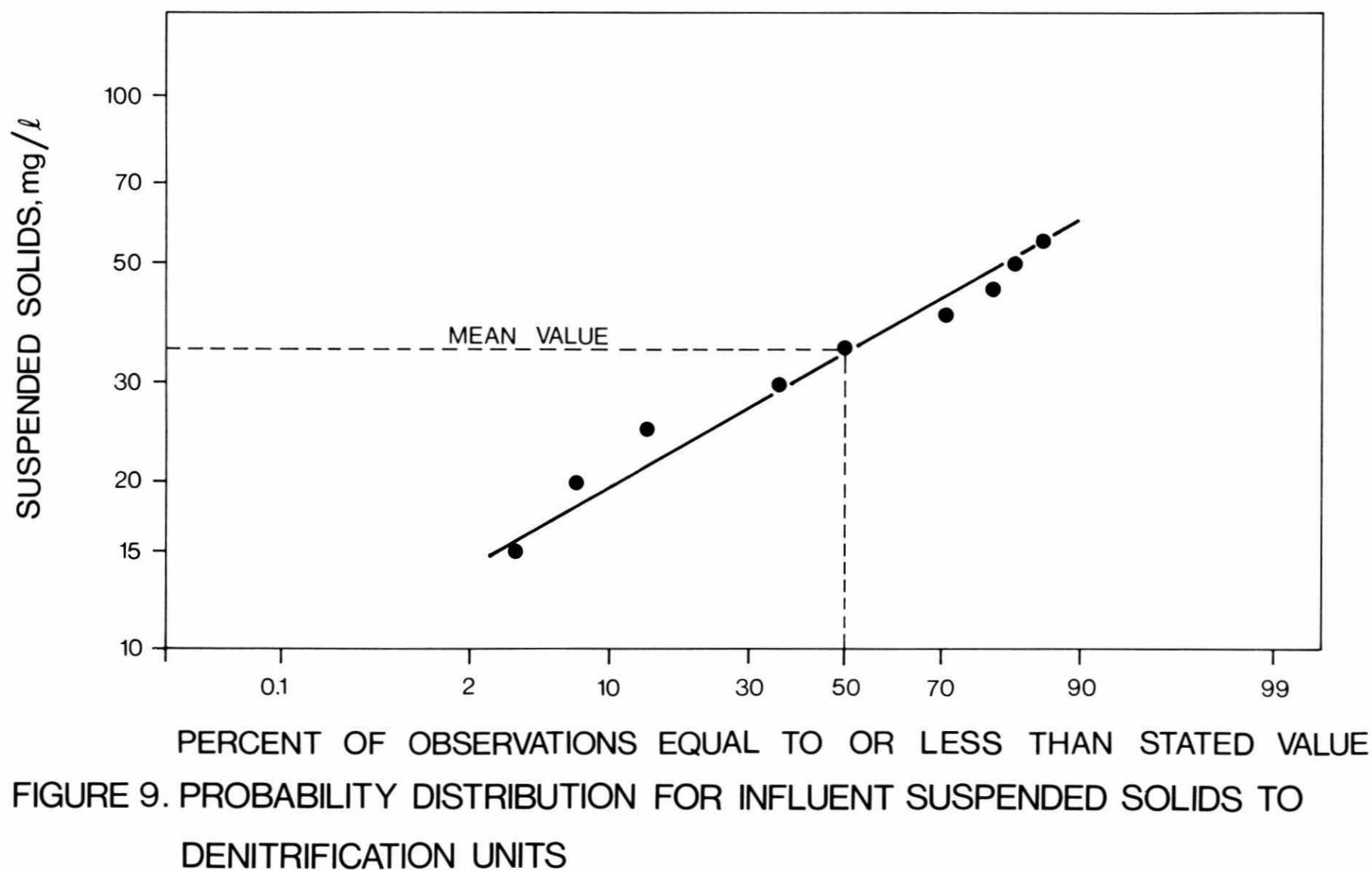
FIGURE 8. PROBABILITY DISTRIBUTION FOR INFLUENT $NO_3 + NO_2 - N$ TO DENI-
TRIFICATION UNITS

FIGURE 9. PROBABILITY DISTRIBUTION FOR INFLUENT SUSPENDED SOLIDS TO DENITRIFICATION UNITS

a result, rates determined from batch stirred reactor studies can be used to predict the continuous results at the same SRT. The Arrhenius expressions found appropriate to describe the three and six day SRT rates are illustrated in Figure 10, together with the nine day SRT models. The activation energy parameter values (E) indicate no apparent difference in the variation of rate with temperature for the continuous results over the SRT range investigated (Table 1). It has been shown statistically that for all practical purposes a common temperature and rate dependency exists for denitrification over a SRT range of three to nine days (Sutton, Murphy, and Jank, 1975). Comparable results derived at other locations (Figure 11) support this contention.

TABLE 1. DENITRIFICATION RATE VARIATION WITH TEMPERATURE IN SUSPENDED GROWTH SYSTEMS

Reactor Configuration	Temp. Range °C	Arrhenius Model Parameters	
		E cal/g-mole	A
Batch			
3 day SRT	6-25	20,400	4.78×10^{14}
6 day SRT	6-25	13,900	4.78×10^{9}
Continuous			
3 day SRT	6-25	12,900	8.76×10^{8}
6 day SRT	6-25	15,900	1.08×10^{11}
9 day SRT	6-25	15,650	8.76×10^{10}
Batch plus Continuous			
3 day SRT	6-25	15,300	5.42×10^{10}
6 day SRT	6-25	15,900	1.15×10^{11}

The rates obtained with the separate sludge system are approximately four times greater than corresponding combined sludge values determined by Barnard (1974).

To optimize the denitrification process, methanol was supplied in excess as an external carbon and energy source. An aerobic stabilization chamber (Figure 7) was operated during part of the experimental
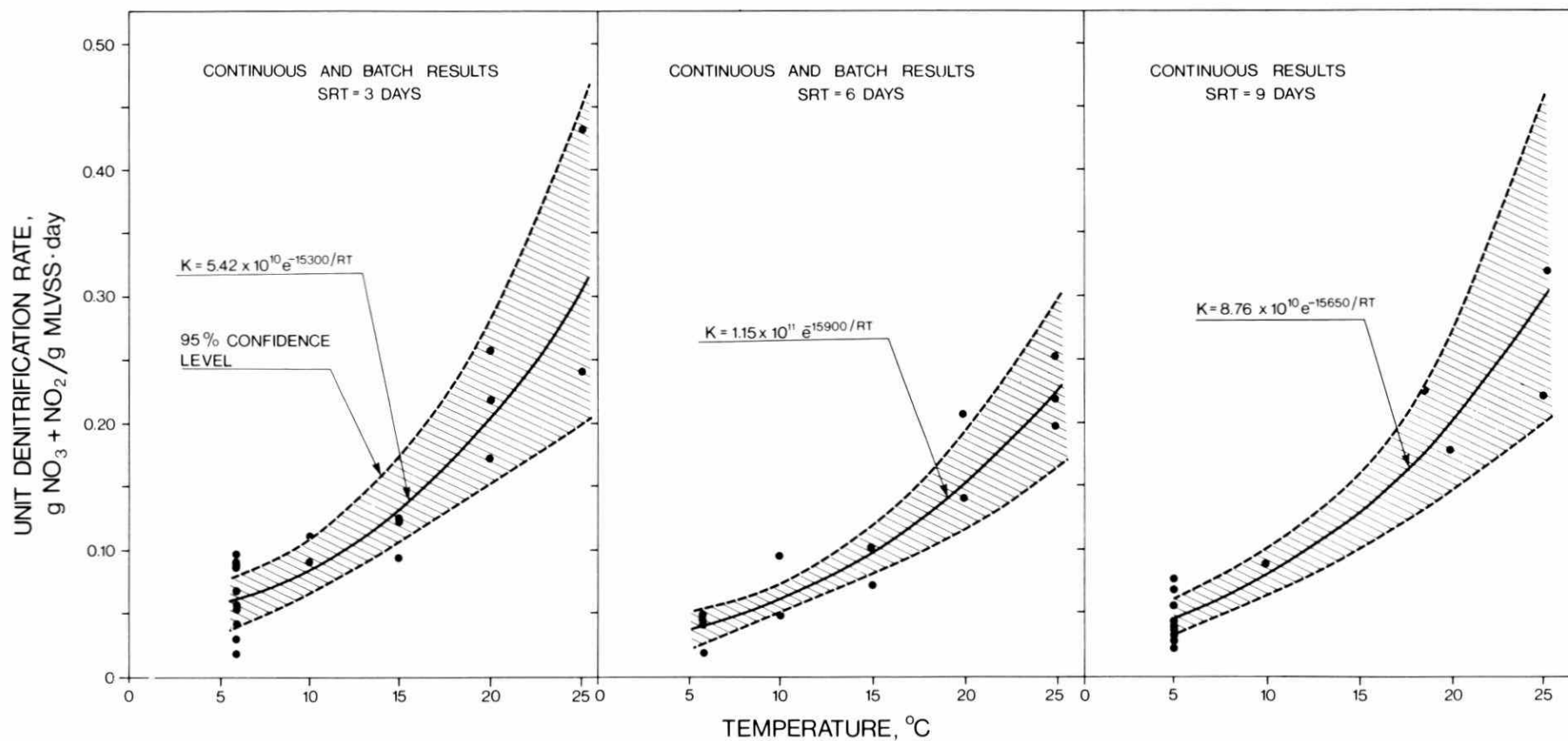
FIGURE 10. TEMPERATURE DEPENDENCY OF DENITRIFICATION IN SUSPENDED GROWTH SYSTEM
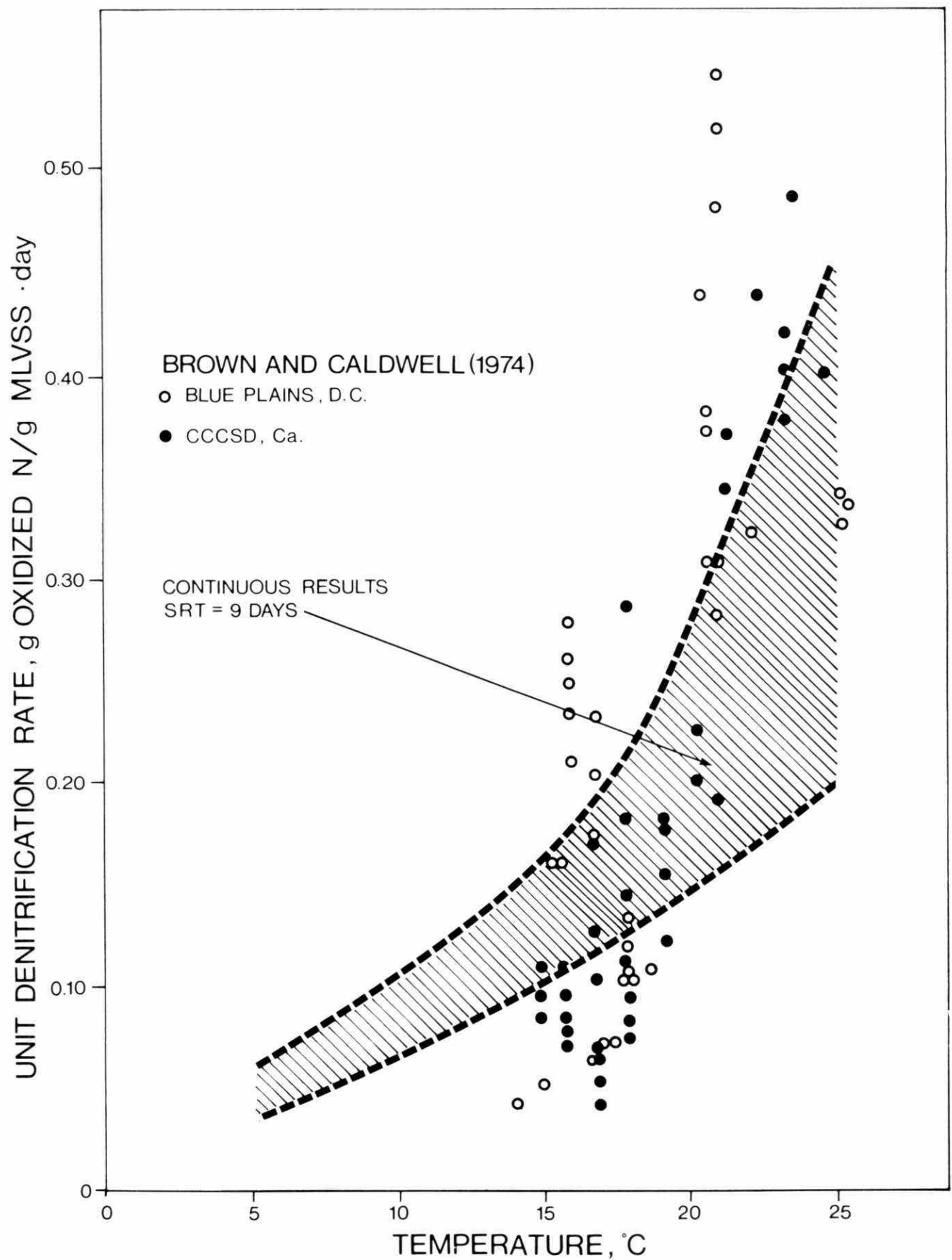
268

FIGURE 11. DENITRIFICATION RATES COMPARED TO RESULTS
REPORTED BY BROWN AND CALDWELL (1974)

period to reduce the methanol residual resulting from overdosing. Using a chromatographic technique to measure methanol, an effective reduction was observed across the aerated unit. The mean methanol concentration was reduced from 6 mg/l to less than 1 mg/l CH_3OH-C (Figure 12) considerably reducing any chance of secondary pollution. The detention time in the aeration chamber was 30 minutes, considerably less than that utilized by Horstkotte, Niles, Parker, and Caldwell (1974).

The release of nitrogen gas from solution following de-nitrification can affect sludge settleability. The aerobic stabilization chamber reduced the mean clarifier effluent suspended solids concentration from 41 to 25 mg/l (Figure 13). No decrease in the rate of denitrification was observed when the stabilization chamber was incorporated into the process.

If a total nitrogen limit is to be met, the nitrogen content of the suspended solids may contribute substantially. With complete nitrification-denitrification and 30 minutes aerobic stabilization, a mean effluent concentration of 1.2 mg/l total Kjeldahl nitrogen could be anticipated (Figure 14).

SUPPORTED GROWTH SYSTEMS

The column reactors investigated through several months of pilot plant operation in this study (Figure 7) can be categorized as fixed bed upflow reactors with porous media, exhibiting minor pressure gradients. The columns containing Pall rings never developed pressure drops above the static heads of the reactors, while limited increases in pressure for the Intalox saddle columns necessitated occasional flushing.

In order to develop a nitrate removal rate expression for the column systems, knowledge of the hydraulic flow pattern is required. Following start-up of the columns, the hydraulic pattern changed from essentially one of plug flow to a situation where a great deal of short circuiting was occurring (Figure 15). This non-ideal flow
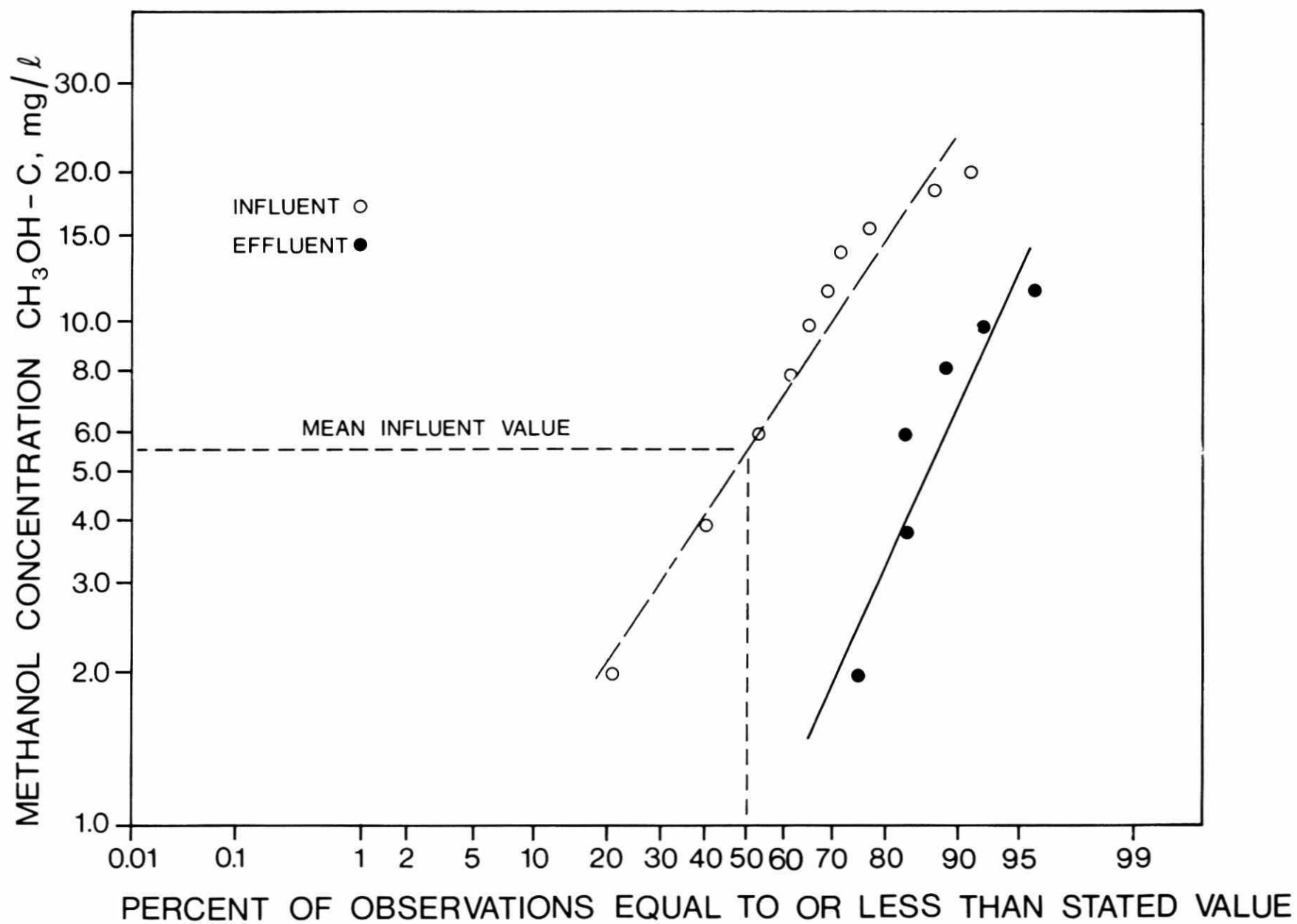
FIGURE 12. METHANOL REDUCTION IN THE AEROBIC STABILIZATION CHAMBER

FIGURE 13. EFFECT OF AEROBIC STABILIZATION ON CLARIFIER EFFLUENT SUSPENDED SOLIDS

FIGURE 14. PROBABILITY DISTRIBUTION FOR NON–FILTERABLE TKN
FROM NITRIFICATION–DENITRIFICATION SYSTEM

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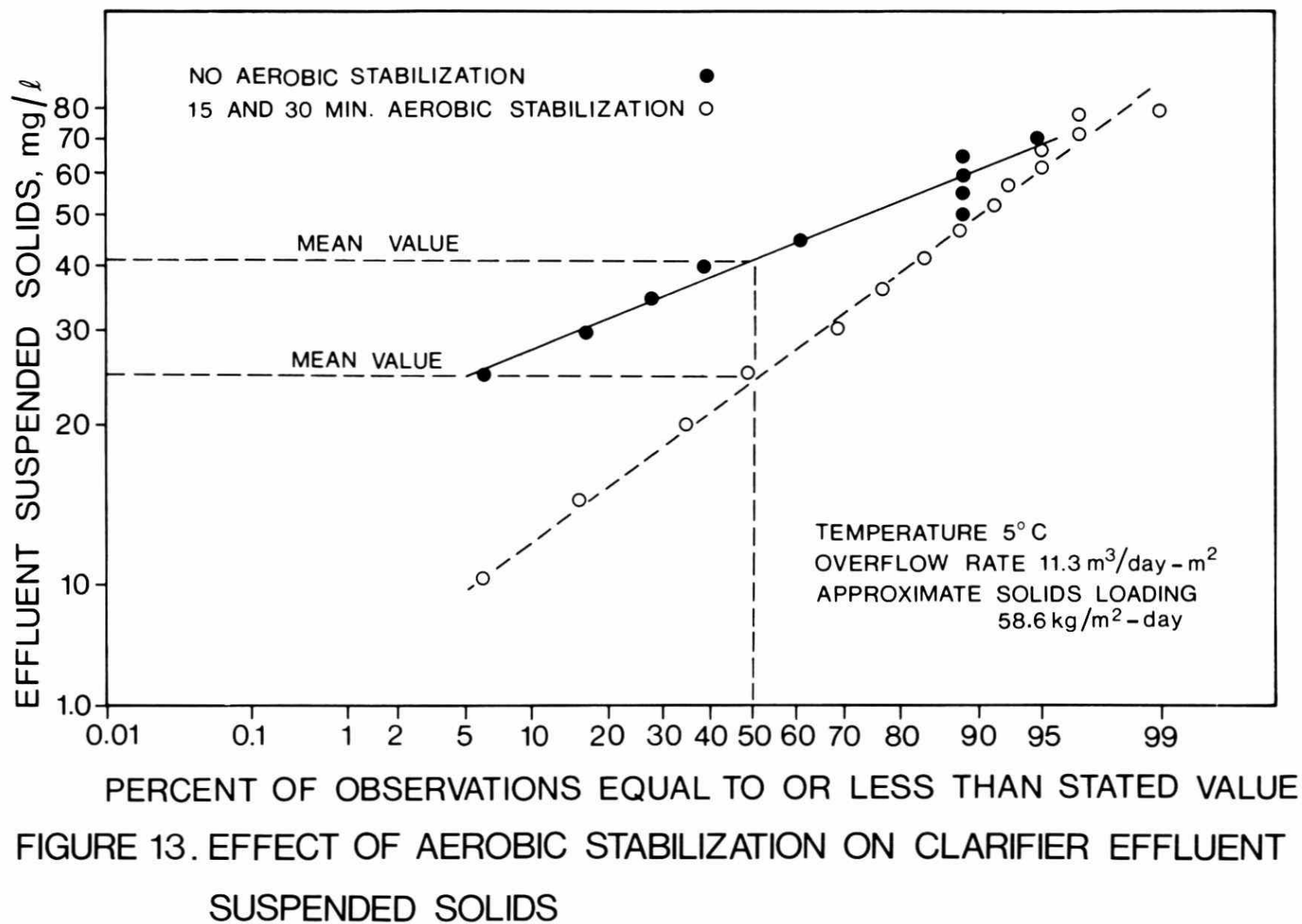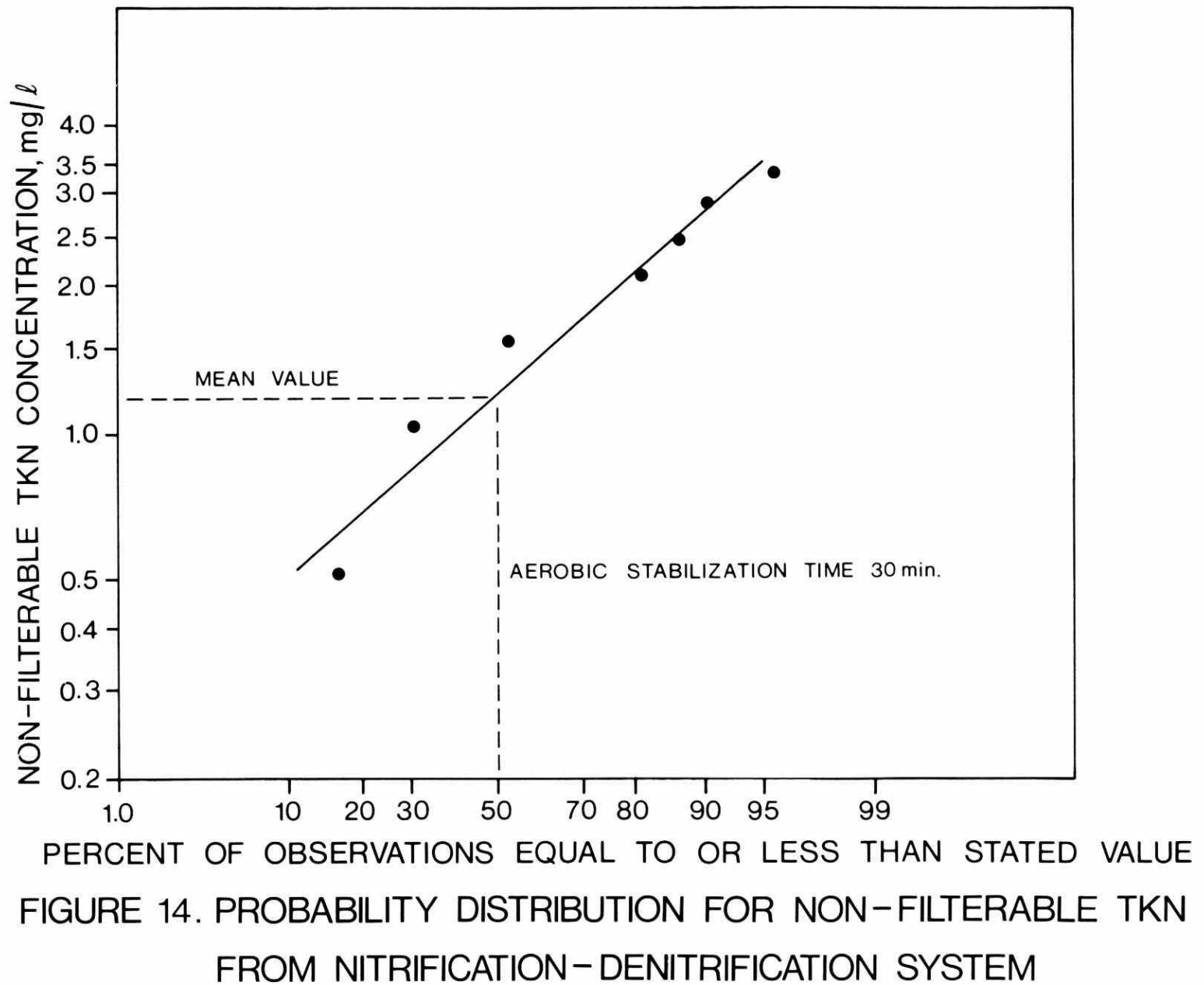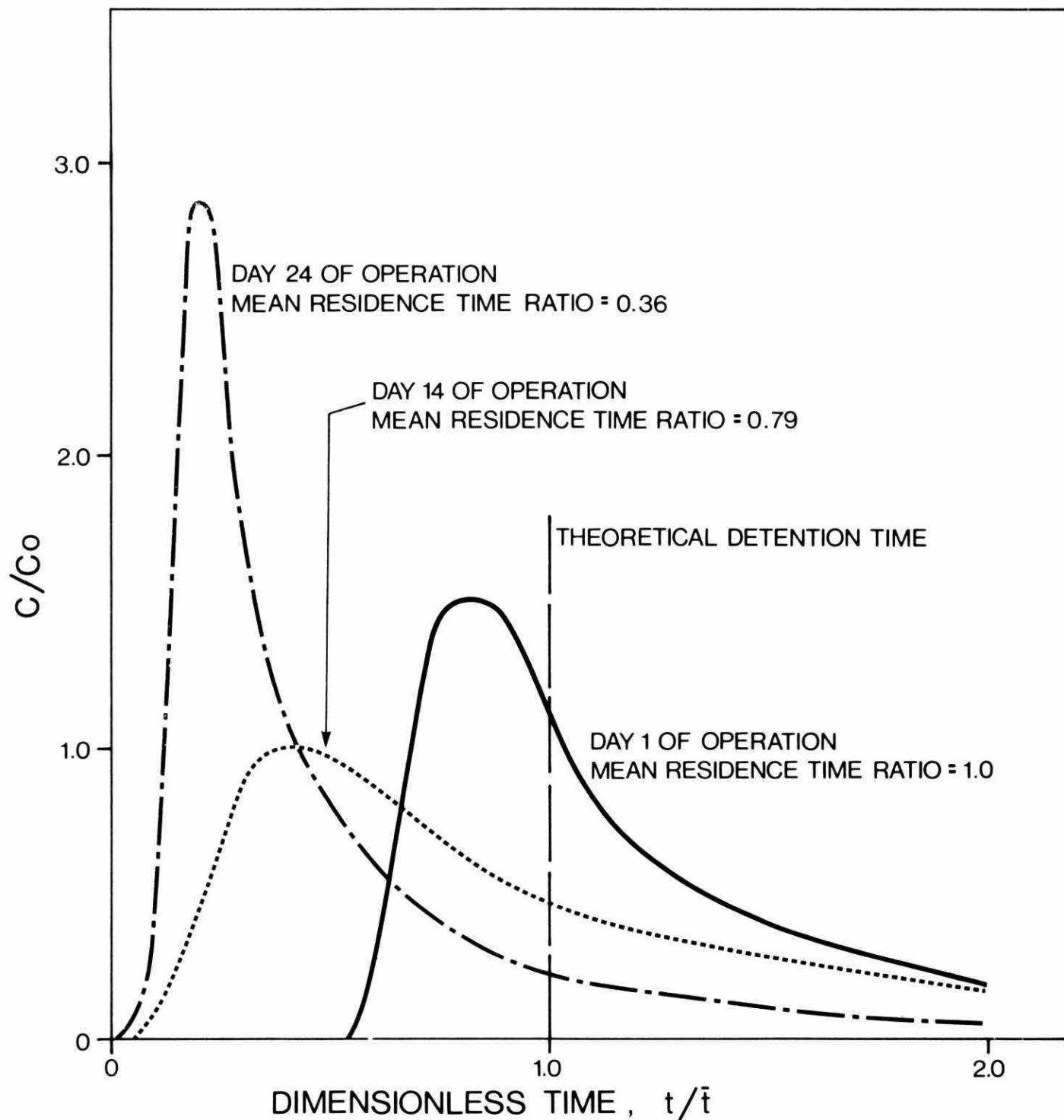
FIGURE 15. PALL RING PACKED COLUMN FLOW CHARACTERIS-
TICS (2.54 cm)

condition was characteristic of the Pall ring and Intalox saddle packed columns. In an attempt to restore the ideal hydraulic conditions and to reduce the minor pressure gradients in the Intalox saddle packed columns, occasional flushing was necessary. This practice reduced the pressure gradients but short circuiting remained a problem in both the Intalox saddle and Pall ring packed columns (Figure 16).

The non-steady hydraulic condition complicates the development of a mechanistic model to describe nitrate removal in the column reactors. A simple zero-order kinetic model was found appropriate to describe the rate variation with nitrate concentration (Sutton, Murphy, and Dawson, 1974). Plotting the surface nitrate removal against theoretical detention time, based on packing porosity, a surface removal rate was determined over a temperature range of 5° to $25^{\circ}C$ for the Intalox saddle packed columns. This procedure is illustrated in Figure 17 for the $20^{\circ}C$ expreimental conditions. For the Pall ring packed columns a dependence on surface area was not found and removal rates were calculated based on volumetric removal (Wilson, 1975).

To describe the temperature dependency of the packed column removal rates, the Arrhenius model was fitted to the data. Unlike the rate data from the other nitrification and denitrification systems the rates determined for the packed columns at each temperature did not have equal variance. Consequently, a linear least squares fit of the removal rates to a linearized form of the Arrhenius equation was not valid. A weighted non-linear least squares technique was used and the weighting factor used was the estimated variance of the removal rate at each temperature. The resulting Arrhenius models (Figures 18 and 19) show a large removal rate variability for both Intalox saddle and Pall ring packed column reactors. The value of the activation energy (E) for the packed column models suggests a reduced temperature sensitivity in comparison to the suspended growth denitrification system (Figure 20).

One advantage of packed column reactors often cited is that a final solids separation step is not required. Although this may be true for packed column reactors exhibiting significant pressure

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BEFORE FLUSHING MEAN
RESIDENCE TIME RATIO = 0.34

AFTER FLUSHING MEAN
RESIDENCE TIME RATIO = 0.32

DIMENSIONLESS TIME, $t/\bar{t}$
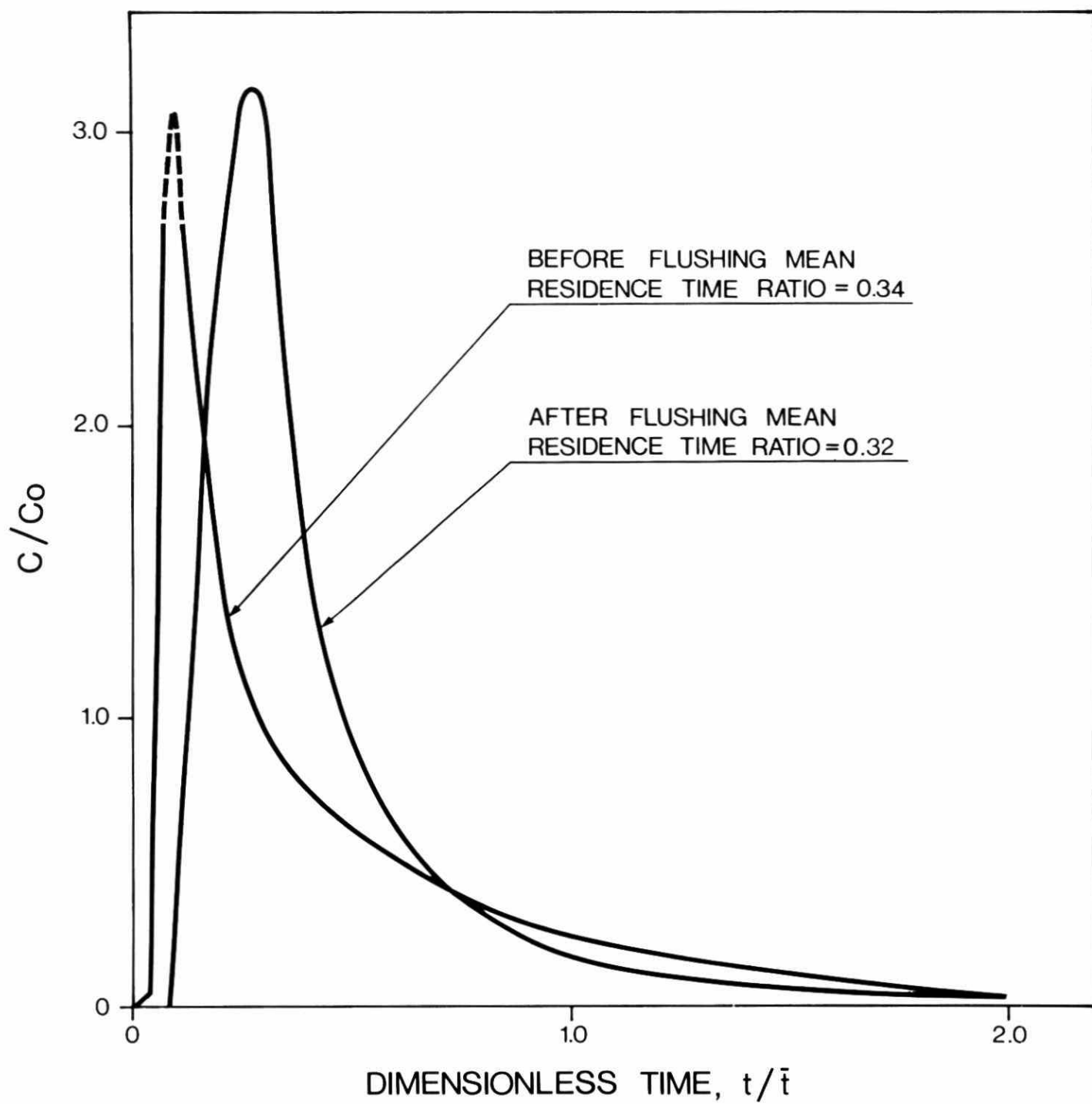
FIGURE 16. EFFECT OF FLUSHING ON PALL RING PACKED
COLUMN FLOW CHARACTERISTICS (2.54 cm)
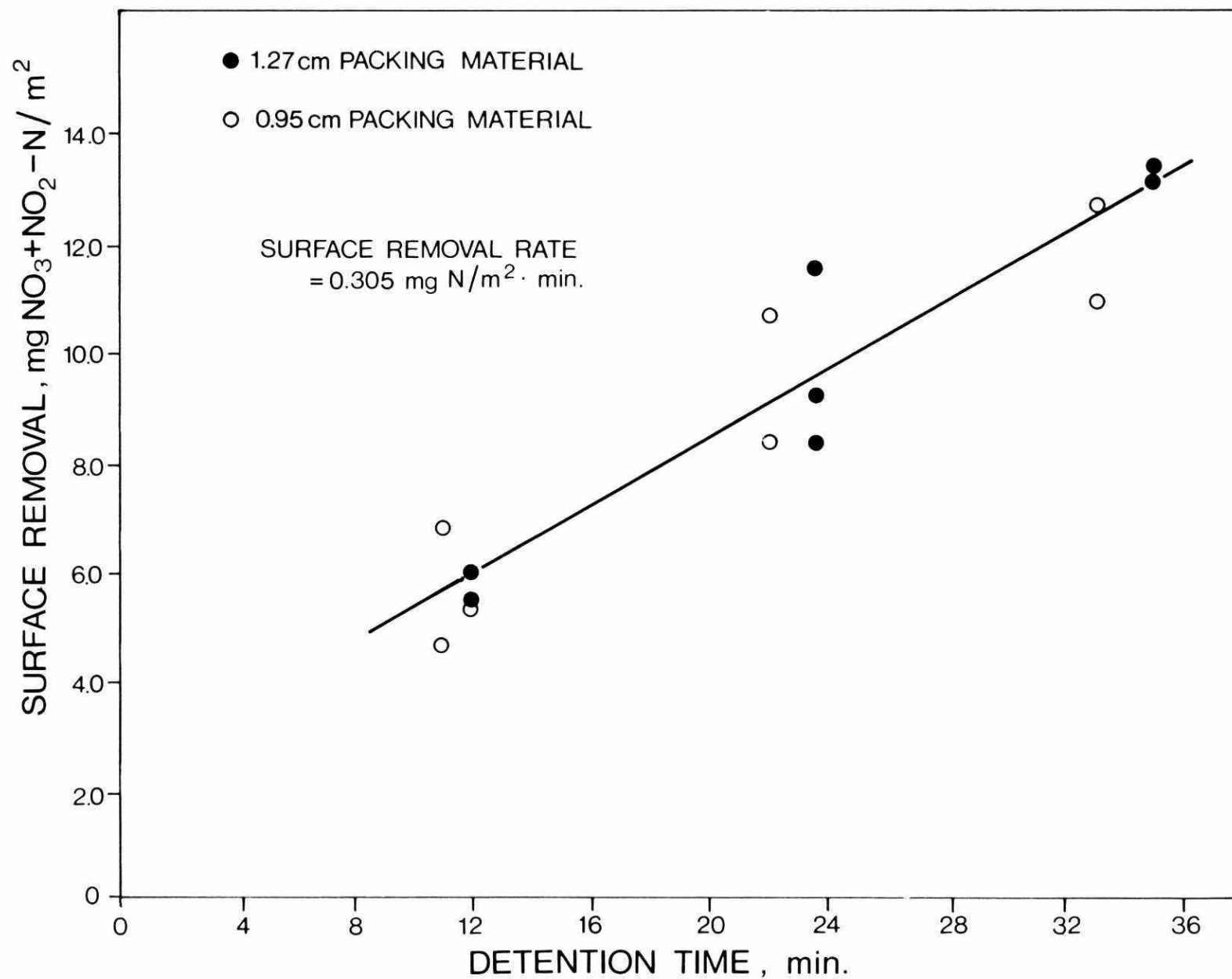
FIGURE 17. DENITRIFICATION RATE DETERMINATION IN INTALOX PACKED COLUMNS
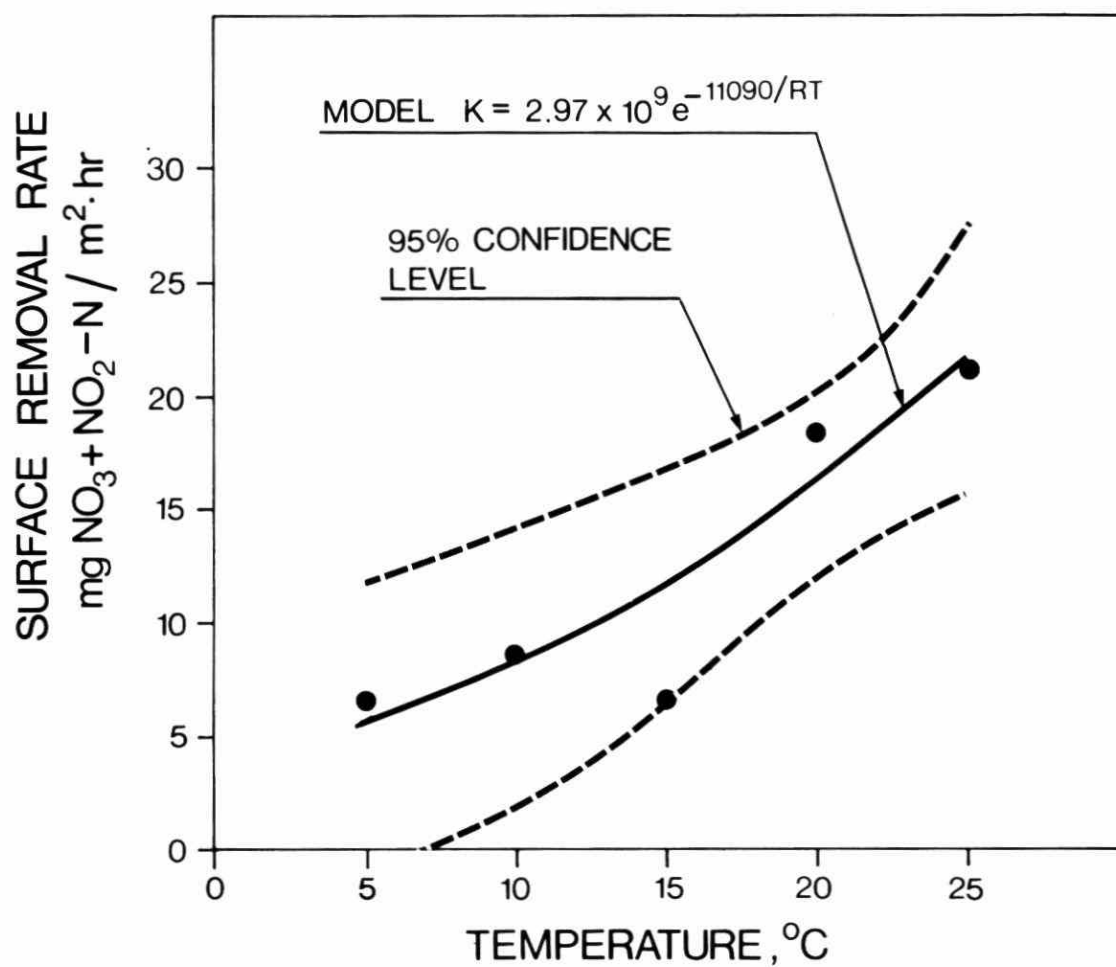
FIGURE 18. TEMPEPATURE DEPENDENCY OF DENITRI-
FICATION IN INTALOX SADDLE PACKED
COLUMNS
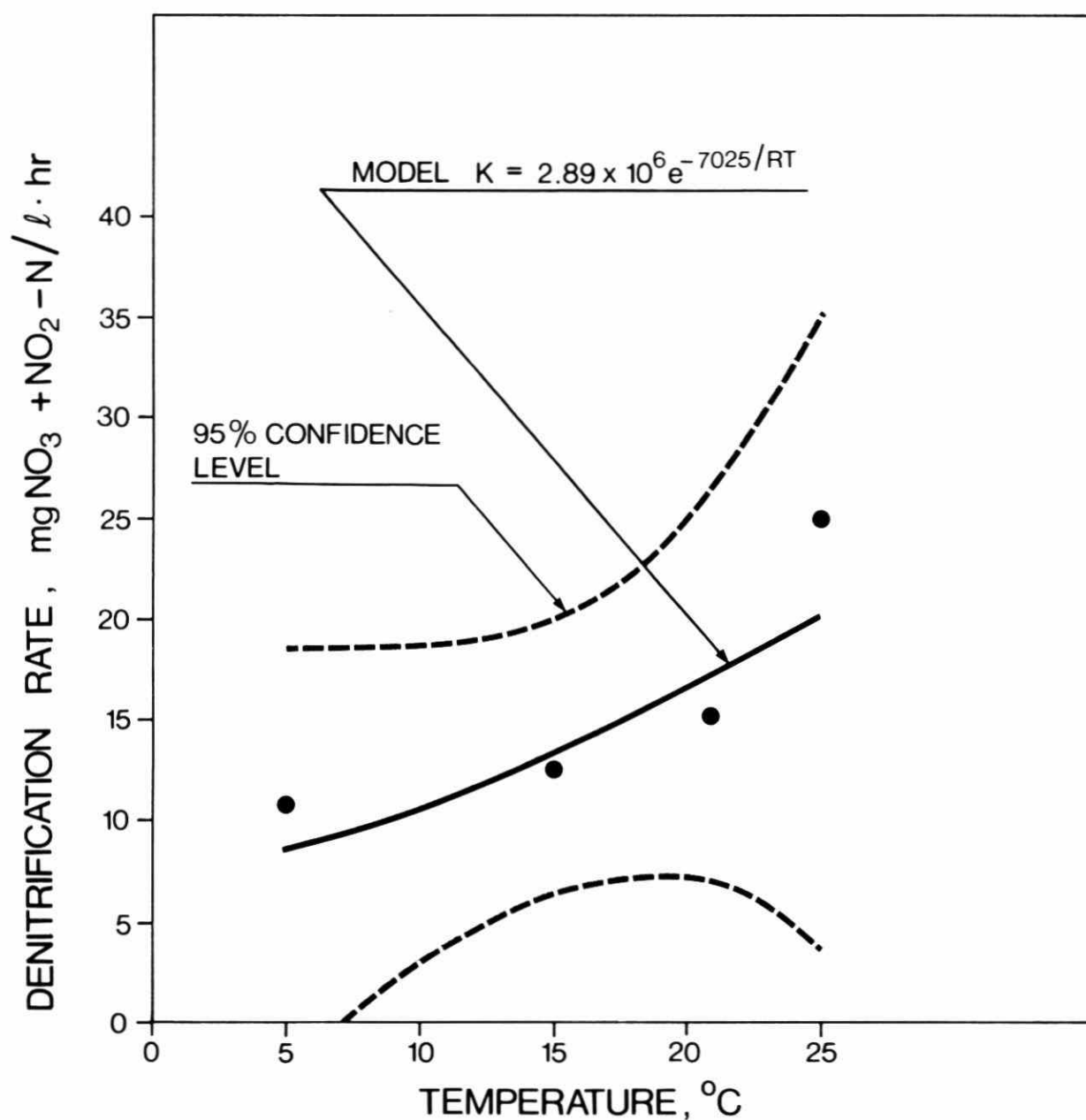
FIGURE 19. TEMPERATURE DEPENDENCY OF DENITRIFICA-
TION IN PALL RING PACKED COLUMNS
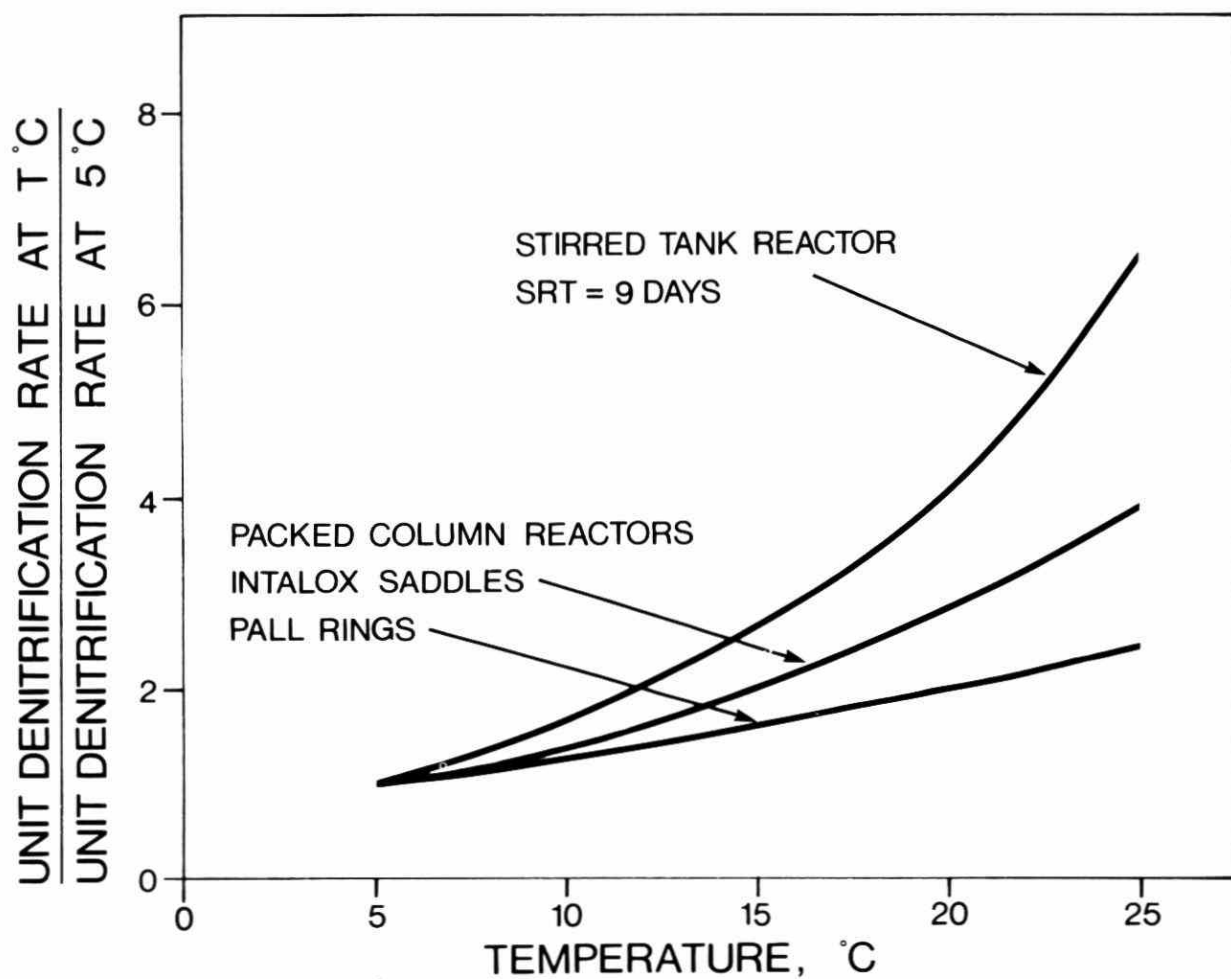
FIGURE 20. DENITRIFICATION RATE DEPENDENCY ON TEM-
PERATURE IN SUSPENDED AND SUPPORTED
GROWTH SYSTEMS

gradients it does not appear to be true for the columns investigated in this study (Figure 21).

A submerged RBC is presently being operated as a denitrification process. This system is identical to the carbon removal-nitrification RBC except that the rotating discs are submerged below the liquid level to help maintain an anaerobic condition for the denitrifying microorganisms. Preliminary results indicate that significant nitrate plus nitrite removals are possible (Figure 22).

A volumetric efficiency comparison of the suspended and supported growth systems indicates an advantage for the suspended growth system (Table 2).

TABLE 2. VOLUMETRIC REMOVAL EFFICIENCY IN SUSPENDED AND SUPPORTED GROWTH DENITRIFICATION SYSTEMS AT $20^{\circ}C$

Denitrification System	Volumetric Removal g $NO_3 + NO_2-$ N/m^3·day	Equivalent MLVSS mg/l
Suspended Growth		
SRT = 9 days (MLVSS = 2500 mg/l)	466	----
Supported Growth*		
Packed Columns:		
Pall Rings		
2.54 cm	398	2140
5.08 cm	398	2140
Intalox Saddles		
0.95 cm	300	1612
1.27 cm	238	1277

*Removal based on volume occupied by the media

DENITRIFICATION DESIGN CONSIDERATIONS

As a final biological nitrogen removal step, suspended or supported growth denitrification systems can be designed for treatment
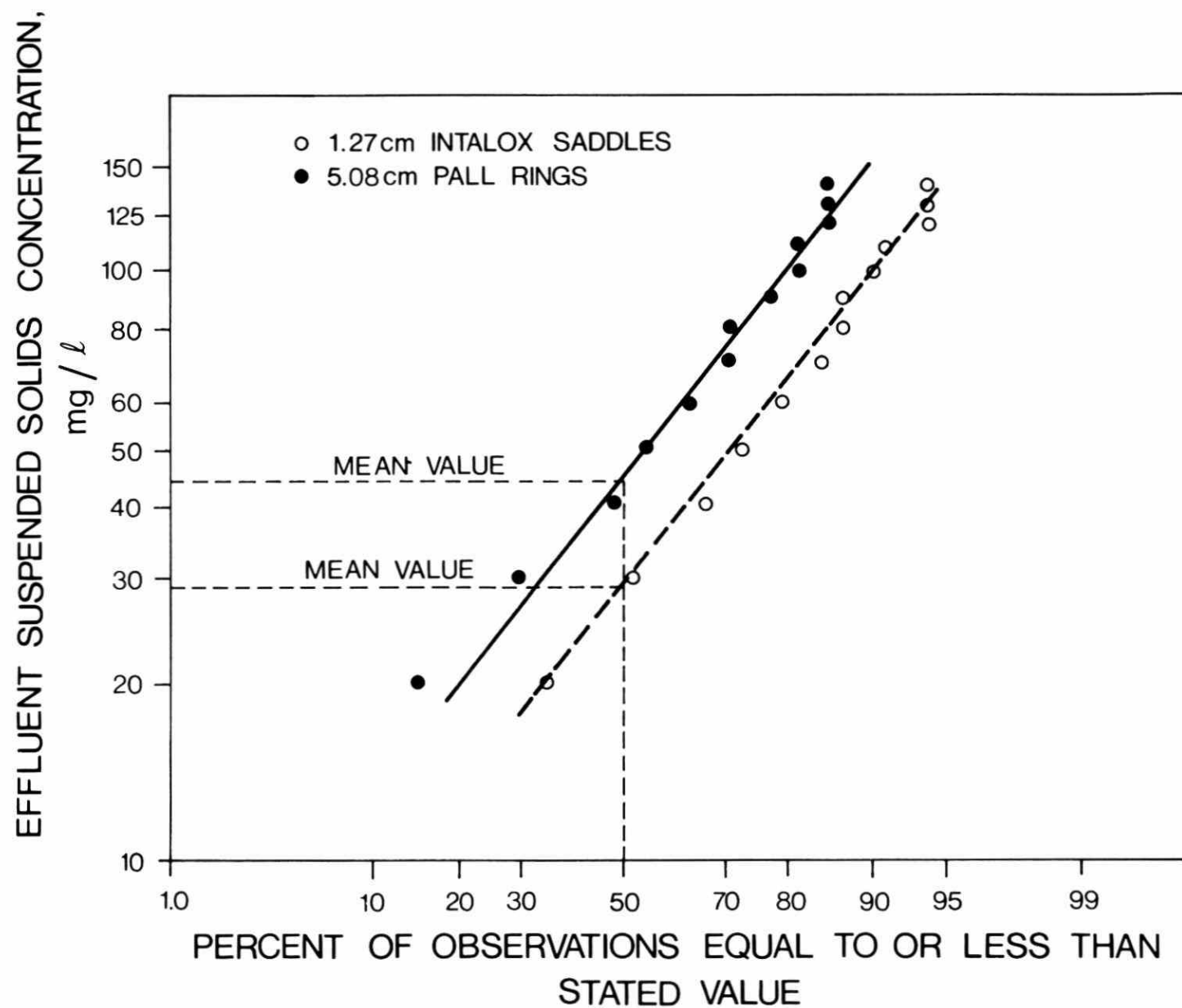
FIGURE 21. EFFLUENT SUSPENDED SOLIDS FROM DENITRIFICATION
PACKED COLUMNS

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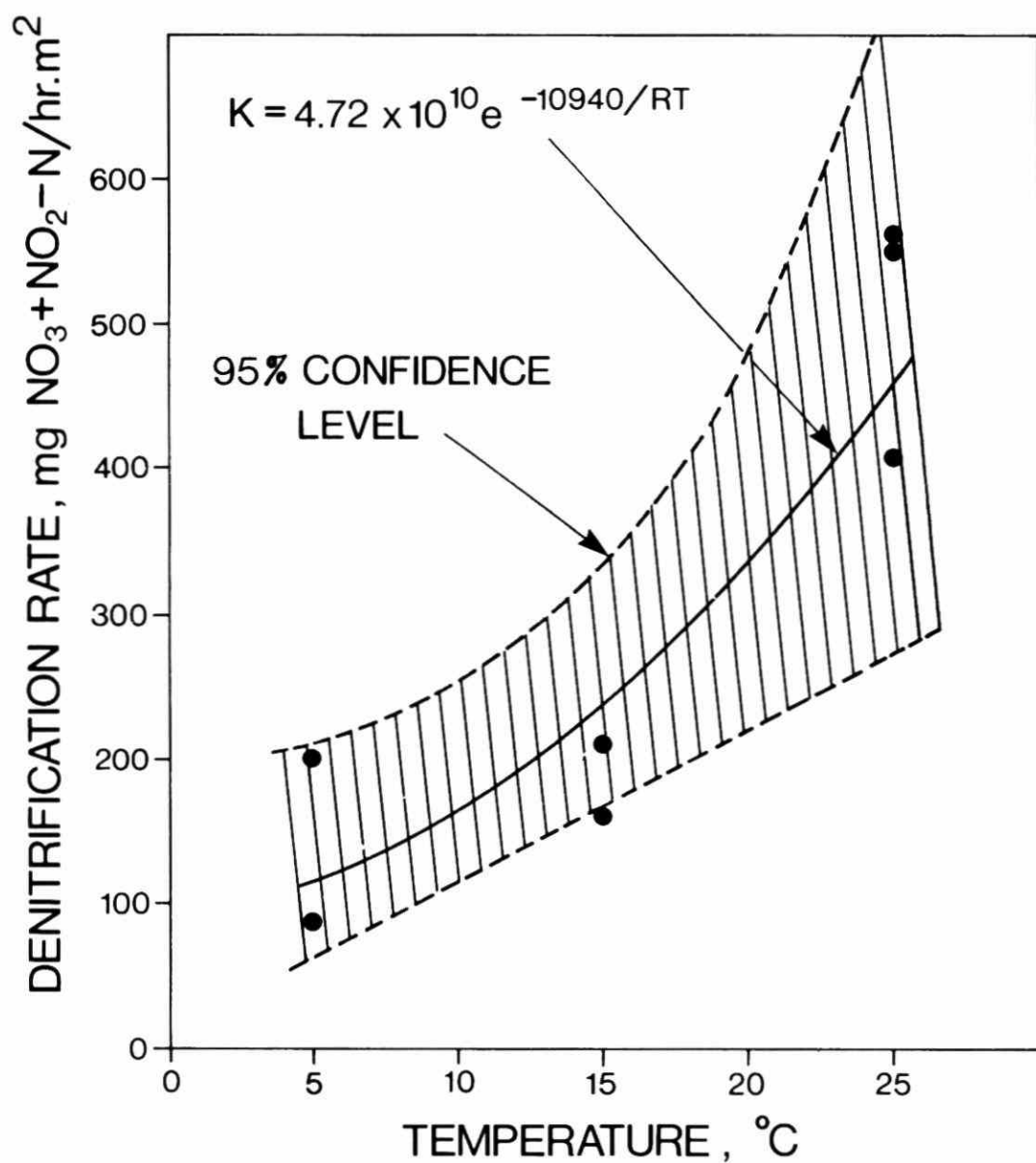
FIGURE 22. TEMPERATURE DEPENDENCY OF DE-
NITRIFICATION IN THE RBC

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of nitrified municipal wastewater effluents. Removal rates will be affected by the temperature, pH, and dissolved oxygen levels in the process reactor.

In designing suspended or supported growth denitrification systems, specific considerations are required. These considerations apply to systems receiving an external carbon source to maximize the denitrification rate.

Suspended Growth Systems

1. For levels of nitrogen commonly found in nitrified domestic waste, the rate of denitrification when expressed as the nitrate plus nitrite removed per unit mass of activated sludge is independent of the concentration of NO_3+NO_2-N.

2. Temperature sensitivity of the denitrification process is essentially independent of solids retention time over the range of three to nine days.

3. The incorporation of an aerobic stabilization chamber with a retention time of 30 minutes will reduce leakage of residual methanol from the process without affecting the denitrification rate. The settleability of the denitrifying sludge is improved.

4. Nitrogen contributed by effluent suspended solids resulted in a mean effluent total Kjeldahl nitrogen concentration of 1.2 mg/l. To consistently attain a total nitrogen effluent requirement of 1 mg/l, some form of additional suspended solids removal will be required.

Supported Growth Systems

1. Fixed bed upflow packed column reactors developing minor pressure gradients will exhibit non-steady hydraulic behaviour resulting in variable denitrification rates. Although variable, significant quantities of nitrate can be removed.

2. For packed columns of the type investigated in this study some form of additional effluent suspended solids removal would be required.

3.   Supported growth denitrification systems appear to be less
          temperature sensitive than suspended growth systems.

ACKNOWLEDGEMENTS

REFERENCES

Barnard, J.L., "Cut P and N without Chemicals", Wat. & Wastes Eng. 7, 33 (1974).

Brown and Caldwell, "Case Histories of Nitrification and Denitrification Facilities", EPA Technology Transfer Design Seminar For Wastewater Treatment Facilities, Orlando, Florida (1974).

Brown and Caldwell, "Process Design Manual for Nitrogen Control", U.S. EPA, Office of Technology Transfer, Washington, D.C. (1975).

Chang, J.P. and Morris, J.G., "Studies on the Utilization of Nitrate by Micrococcus denitrificans", J. gen. Microbiol., 29, 301-310 (1962).

Christensen, M.H. and Harremoes, P., "A Literature Review of Biological Denitrification of Sewage", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Christensen, M.H., Harremoes, P. and Jensen, O.R., "Combined Sludge Denitrification of Sewage Utilizing Internal Carbon Sources", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Dawson, R.N. and Murphy, K.L., "Factors Affecting Biological Denitrification of Wastewater", <u>Advances in Water Pollution Research</u>, edited by S.H. Jenkins, Pergamon Press, Oxford, England (1973).

Delwiche, C.C., "Denitrification", <u>A Symposium on Inorganic Nitrogen Metabolism</u>, edited by McElroy & Glass, John Hopkins Press, Baltimore, Maryland, p. 233 (1956).

English, J.N., Carry, C.W., Masse, A.M., Pitkin, J.B. and Dryden, F.D., "Denitrification in Granular Carbon and Sand Columns", <u>J. Wat. Pollut. Control Fed.</u>, <u>46</u>, 1, 28 (1974).

Harremoes, P., "The Significance of Pore Diffusion to Filter Denitrification", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Horstkotte, G.A., Niles, D.G., Parker, D.S. and Caldwell, D.H., "Full Scale Testing of a Water Reclamation System", <u>J. Wat. Pollut. Control Fed.</u>, <u>46</u>, 1, 181 (1974).

Jeris, J., Beer, C. and Mueller, J.A., "High Rate Biological Denitrification Using a Granular Fluidized Bed", <u>J. Wat. Pollut. Control Fed.</u>, <u>46</u>, 9, 2118 (1974).

Ludzack, F.J. and Ettinger, M.B., "Controlling Operation to Minimize Activated Sludge Effluent Nitrogen", <u>J. Wat. Pollut. Control Fed.</u>, <u>34</u>, 9 920 (1962).

McCarty P.L., "Nitrification-Denitrification by Biological Treatment", Correspo Jence Conference on Denitrification of Municipal Wastes, University of Massachusetts, Mass. (1973).

McCarty, P.L., Beck, L. and St. Amant, P., "Biological Denitrification of Wastewaters by Addition of Organic Materials", Proceedings 24th Industrial Waste Conf., 1271, Purdue University, Indiana (1969).

Matsché, N.F. and Spatzierer, G., "Austrian Plant Knocks Out Nitrogen", Water and Waste Eng., 12, 1, 18 (1975).

Moore, S.F. and Schroeder, E.D., "An Investigation of the Effects of Residence Time on Anaerobic Bacterial Denitrification", Water Research, 4, 685 (1970).

Mulbarger, M.C., "Nitrification and Denitrification in Activated Sludge Systems", J. Wat. Pollut. Control Fed., 43, 10, 2040 (1971).

Murphy, K.L. and Sutton, P.M., "Pilot Scale Studies on Biological Denitrification", Presented at the 7th International Conference on Water Pollution Research, Paris, France (1974).

Myers, J. and Matsen, F.A., "Kinetic Characteristics of Warburg Manometry", Arch. Biochem. & Biophys., 55, 373 (1955).

Painter, H.A., "Microbial Transformations of Inorganic Nitrogen", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Requa, D.A. and Schroeder, E.D., "Kinetics of Packed Bed Denitrification", J. Wat. Pollut. Control Fed., 45, 8, 1696 (1973).

Schmidt, B. and Kampf, W.D., "Über den Einfluss des Sauerstoffs auf die Denitrifikationsleistung von Pseudomonas fluorescens", Arch. Hyg. Bakt., 146, 171 (1962).

Stensel, H.D., Loehr, R.C. and Lawrence, A.W., "Biological Kinetics of Suspended Growth Denitrification", J. Wat. Pollut. Control Fed., 43, 2, 249 (1973).

Sutton, P.M., Murphy, K.L. and Dawson, R.N., "Continuous Biological Denitrification of Wastewater", Technology Development Report, EPS 4-WP-74-6, Environment Canada (1974).

Sutton, P.M., Murphy, K.L., and Jank, B.E., "Nitrogen Control: A Basis for Design with Activated Sludge Systems", IAWPR, Conference on Nitrogen as a Water Pollutant, Proceedings VOL. 3, Copenhagen, Denmark (1975).

Sutton, P.M., Murphy, K.L., and Dawson, R.N., "Low Temperature Biological Denitrification of Wastewater", J. Wat. Pollut. Control Fed., 47, 1, 122 (1975).

Sutton, P.M. and Jank, B.E., "Design Considerations for Biological Nitrification-Denitrification Systems", EPS Technology Transfer Seminar on Alternatives for Nutrient Control, Kelowna, B.C. (1975).

Wilson, R.W., Continuous Fixed Film Biological Nitrification and Denitrification of Wastewater, M.Sc. Thesis, McMaster University, Hamilton, Canada (1975).

Wilson, R.W., Murphy, K.L., Sutton, P.M. and Jank, B.E., "Nitrogen Control: Design Considerations for Supported Growth Systems", Presented at 48th Annual Conference of Wat. Pollut. Control Fed., Miami Beach (1975).

Wuhrmann, K., "Nitrogen Removal in Sewage Treatment Processes", Verh. Int. Ver. Limnol., 15, 580 (1964).

NITROGEN REMOVAL
AN EVALUATION OF FULL SCALE BIOLOGICAL SYSTEMS

by

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INTRODUCTION

During the past decade considerable effort has been directed to the removal of nitrogen from sewage effluents via biological means. This has been a result of the international community becoming aware of the need for carbon, nitrogen and phosphorus reduction in sewage effluents to protect the ecology of the often sensitive receiving waters.

Extensive laboratory investigations have contributed much to the knowledge of nitrogen reactions within activated sludge and parallels can be seen between biochemical exchanges that occur in soils and surface waters and the biological sewage treatment process. Since the biological process appears to offer the most economical method of treating sewage, particularly on a large scale, this form was chosen by many for nitrogen removal. Consequently, many full scale evaluations using existing or slightly altered conventional activated sludge plants have been initiated to study nitrification-denitrification efficiencies under varying climatic conditions.

The aim of this report is to review some the the recent accomplishments using the single-stage nitrification-denitrification scheme for nitrogen removal on a full scale basis. Included in this evaluation are studies performed by the Ontario Ministry of the Environment under partial financing from the Canada/Ontario Agreement Arrangement, and a survey of the Ministry's waste treatment facilities outlining present nitrification efficiencies.

For the purpose of classification in this paper, the term "full scale" will apply to those facilities that process 10,000 gallons per day (45,000 l/day) or more of sewage.

REVIEW OF FULL SCALE BIOLOGICAL PROCESS

Wuhrmann (11) concluded in a 1963 report out of studies on oxygen effects on sewage purifcation, that regardless of oxygen concentration in aeration mixed liquor, full conversion of ammonia to nitrate (nitrification) would not take place unless adequate hydraulic detention was offered in the aeration basin. Furthermore, nitrification did proceed well in experiments where the aeration basin dissolved oxygen (DO) was only one mg/l when low organic loadings and high sludge concentrations (above 3000 mg/l) were utilized. Some attention was also focused on the possibility that individual aeration floc size and/or shape may have some bearing on internal floc DO levels and subsequently may affect nitrogen conversion efficiencies.

Periods of anaerobiosis of aeration mixed liquors of up to five hours showed little effect on endogenous bacterial respiration or substrate utilization when returned to the aerobic state. Moreover, during the anoxic sequence nitrate reduction to nitrogen gas (denitrification) took place at a high rate. From these experiments, Wuhrmann concluded that the efficiency of the aeration mixed liquor to metabolize organic materials is not deterred after a period of anoxia.

Continuing studies by Wuhrmann (10) involved the application of a 13,300 gal/day (60.5 m^3/day) activated sludge plant for nitrification-denitrification followed by clarification and clarifier sludge return to the nitrification basin influent (Figure 1).

Total nitrogen removals utilizing this process arrangement during one year's observation ranged from 35% in winter to 90% in the summer months, without the use of an external carbon addition to the denitrification reaction basin. The hydraulic detention times (2 hours) and solids concentration were augmented during low liquid temperature periods, but nitrification and denitrification still did not proceed fully. A minimum of four days sludge age was required to maintain nitrification at $10-12^{\circ}C$ sewage temperature.

Bishop (2) reported later on a process variation of the single-sludge system in which two chambers were operated alternately as aerobic and anoxic reaction basins for the nitrification-denitrification process. These studies were performed over a period of nine months on a plant designed for 50,000 lgpd (189 m^3/day) sewage flow (Figure 2).

(Wuhrmann 1965)

$$NH_3^+ \xrightarrow{} NO_2^- \xrightarrow{} NO_3^-$$

$$\begin{array}{c} NO_2 \\ NO_3 \end{array} \xrightarrow{} N_2, \; N_2O \uparrow$$

Primary
Effluent → [Aeration] → [Denitrification] → (Sec. Clarifier) →
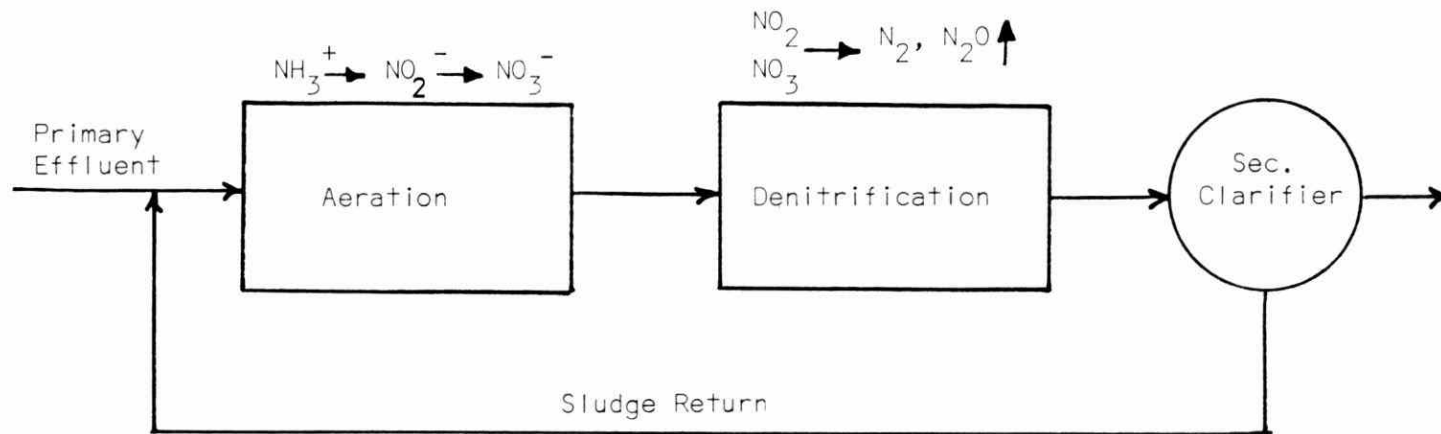
Sludge Return

Optimum Conditions and Results

    Basin detention - aeration 1.7 hr
                       - denitrification 2.8 hr

    Basin S. Solids - aeration and denitrification 5500 mg/l

    Sewage Retention Time (SRT) - 2-3 days

    Sec. Sludge Return Rate   - 100% of flow

    Sewage Temperature   - 18°C

Sec. Effluent Total Nitrogen   - 3-6 mg/l

Sec. Effluent Nitrate Nitrogen - 2-3 mg/l

Figure 1

Bishop (1974)
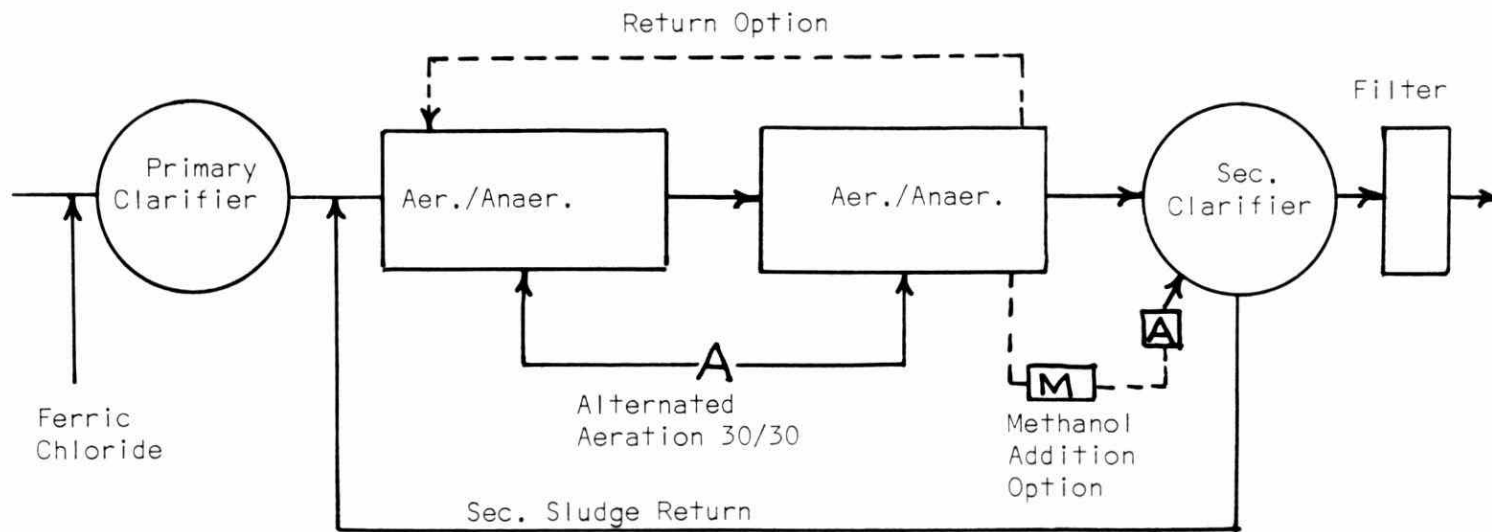
Optimum Conditions and Results

Basin Detention	- aeration/denitrification 6-8 hr
Basin S. Solids	- 3000-4000 mg/l
F/M wt. BOD/wt. VSS	- 0.01-0.1
SRT	- 15-20 days
Sewage Temperature	- 15°C
Sec. Effluent Total Nitrogen	- 3.8-8.6 mg/l
Sec. Effluent Nitrate Nitrogen	- 2.3-4.4 mg/l
Sec. Effluent Ammonia Nitrogen	- 0.3-2.4 mg/l

Figure 2

The procedure involved thirty minute cycles and two aeration
basins were alternately aerated and then mixed without air. Methanol
was added to the process mixed liquor in the chamber between the last
aeration basin and the secondary clarifier as an option when residual
nitrate concentrations escaping the main aeration basins were too high.
A conventional sludge recycle from the secondary clarifier was used
and on occasions an internal aeration recycle was utilized.

Ferric chloride was added for phosphorus removal at a dosage
of 45 mg/l to the primary clarifier flow at one stage of the study.
Unfortunately, this treatment lowered the chemical oxygen demand-to-total
Kjeldahl nitrogen ratio (COD:TKN) resulting in a decrease of 20% in nitrogen
removal efficiency. Aluminum sulphate at a dosage of 20 mg/l was added
prior to tertiary filtration for effective phosphorus removal during two
months of the project.

During the winter months, a bulking mixed liquor developed in
the aeration basins and resultant high sludge volume indexes (SVI)
causing secondary clarifier bulking necessitated lowering the sewage
flow to 30,000 lgpd (113 m^3/day). Secondary clarifier overflow rates
were raised from 300 to above 500 lgpd/ft^2 by increasing sewage flow at
the onset of warmer liquid temperatures and decreased SVI's in summer.
The high SVI's were diagnosed to be a result of overgrowths of filamentous
bacteria thought to be due to very low process food-to-microorganisms
ratios (F:M's lower than 0.1 lb BOD/lb VSS).

Methanol as a carbon source for denitrification was added to
the process only during the summer months. A drop in nitrogen removal
took place with this supplementation and laboratory batch studies confirmed
that methanol caused an inhibition to the nitrification reactions.

Overall nitrogen reduction across the biological process were
75 to 85% without the application of methanol. Nitrification reactions
proceeded well when mixed liquor DO concentrations were above two and
denitrification occurred when the air was turned off and the DO's approached
zero.

At Pretoria South Africa, Barnard (1) studied a 22,000 lmgpd
(100^3/day) plant for nitrification-denitrification called the "Bardenpho"
process in which four basins in series followed by clarification were used

for nitrogen removal. The first and third basins were operated in the anoxic state with the second and fourth being aerated. Sludge return was provided from the secondary clarifier and the second basin effluent back to the biological process influent (Figure 3).

Nitrogen removals exceeding 90% were attained during an 18 month study, but effluent nitrates did rise above 5 mg/l during winter when basin temperatures dropped below 20°C. By increasing the basin hydraulic detention time by one hour, increased denitrification was obtained at liquid temperatures approaching 15°C.

Biological phosphorus removal was observed across the process when denitrification reactions were at their maximum efficiency. This luxury uptake occurred in the fourth basin under aeration when the preceding denitrification basin reactions produced a profuse orthophosphate release.

Studies on a modified extended aeration facility for nitrogen removal by Nicholls (5) produced excellent results on domestic sewage without the use of an external carbon source for denitrification. Two plants of 6 Imgd (27 ml/day) and 18 Imgd (80 ml/day) capacities were studied under a liquid temperature range of 13-15°C. The physical layout of the plant is shown in Figure 4.

The technique of combined nitrification-denitrification was accomplished by turning off four mechanical aerators at the aeration entrance. This created an anoxic zone but maintained enough mixing to keep the mixed liquor suspended solids in motion. This area became the denitrification basin and received raw sewage as well as recirculated secondary clarifier sludge. Following this zone was a large extended aeration process.

Suspended solids concentrations in the total basin varied between 2500 and 6000 mg/l and there were no intervening walls between the various areas. It was shown that as the suspended solids concentration in the basin approached 2500 mg/l denitrification, reactions dropped severely.

It was observed that the resultant low nitrate mixed liquor thickened more efficiently than the nitrified sludge. Furthermore, dewatering tests showed the denitrified sludge to be superior for sludge conditioning.

(Barnard 1973)

Bardenpho Process

Recycle

4:1

Primary Effluent → [2 hr An.] → [6 hr Aer.] → [6 hr An.] → [2 hr Aer.] → (Sec. Clarifier) →

Return 1:1
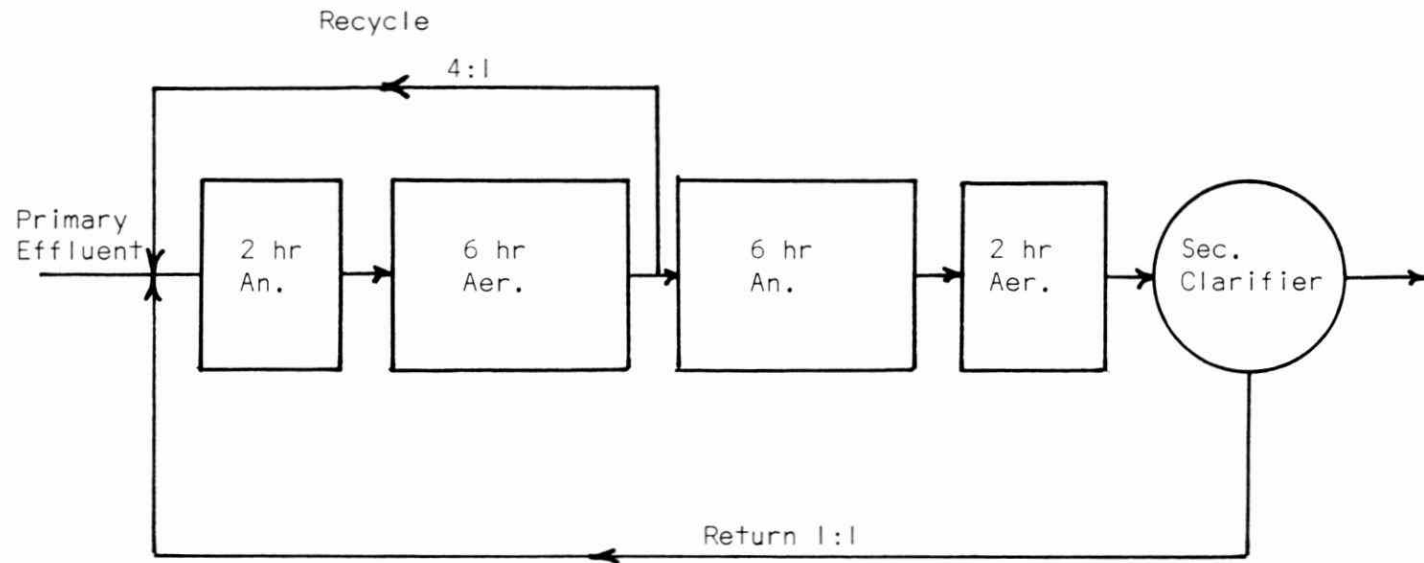
Optimum Conditions and Results

Basin S. Solids	– 4500 to 6000 mg/l
SRT	– 16-18 days (25 days at 6000 mg/l SS)
Sewage Temperature	– 20°C
Sec. Effluent Total Nitrogen	– 2.5 mg/l
Sec. Effluent Nitrate "	– 1.0 mg/l
Sec. Effluent Ammonia "	– 0.3 mg/l

Figure 3

Raw Sewage

Degritter

Anoxic Zone

Aeration Basin 1

Aeration Basin 2

Clarifier

Screens

return
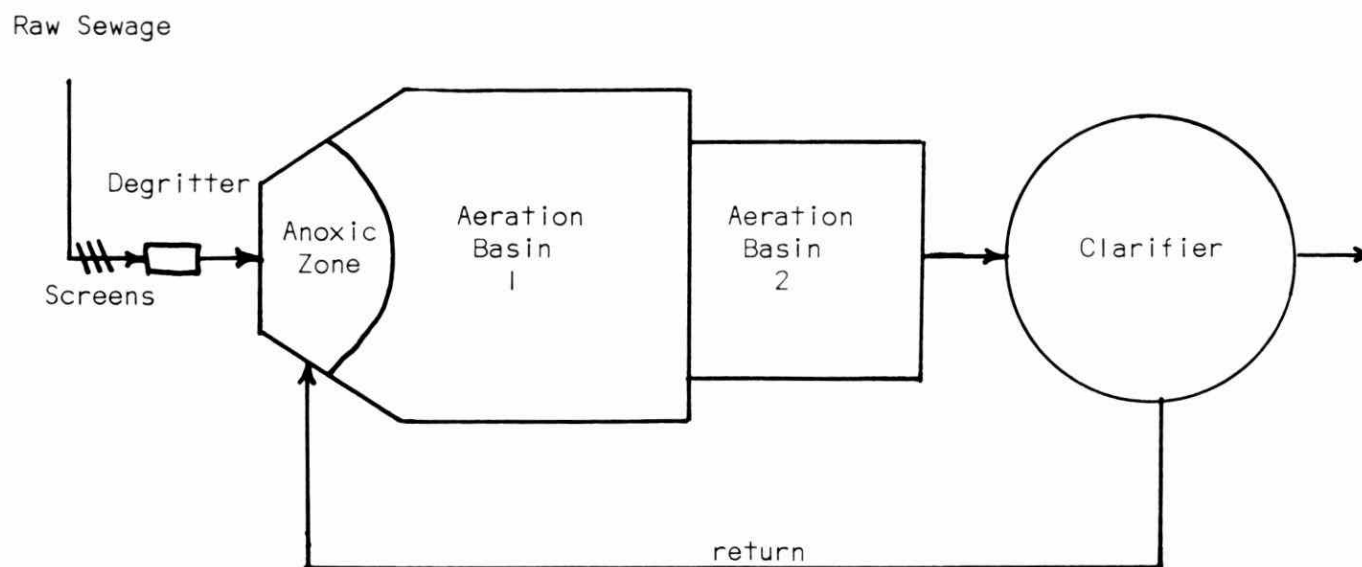
Optimum Conditions and Results

Basin Detention	- 20 hrs total
Basin S. Solids	- 6000 mg/l
SRT	- 30-40 days
Sewage Temperature	- 15°C
Sec. Effluent **Total** Nitrogen	- 3 mg/l
Sec. Effluent Nitrate Nitrogen	- 1 mg/l
Sec. Effluent Ammonia Nitrogen	- 1 mg/l

Figure 4

Nitrate removals of 98% and total nitrogen reductions of 85% were accomplished with process solids retention times (SRT) exceeding 30 days. An aeration basin hydraulic detention time of 20 hours was normal.

Studies at the Vienna, Austria, Blumental waste treatment plant by Matsche (14), entailed the use of a 19.2 Imgd (86.4 ml/day) oxidation ditch for nitrogen removal. Nitrification-denitrification was achieved simultaneously by reducing the number of operating mammoth rotors (aerators) along the length of two aeration channels. By manipulating the number and combination of rotors in operation, aerobic and anoxic pockets were created in various channel locations (Figure 5).

For purposes of this study, two aeration channels were connected in series and three rotors were usually operated in each channel. On occasions a fourth rotor was implemented in the first aeration channel to maintain nitrification at peak process loading periods. With the series flow configuration, the mixed liquor DO levels tended to drop to values below 1 mg/l between the rotors in the first aeration channel and denitrification proceeded rapidly. In the second aeration channel with reduced sewage loading the DO changes were less abrupt and wider areas of less DO depletion were observed by profile sampling.

Oxygen utilization rates of the mixed liquors were normally between 80 and 100 mg O_2/l/hr in the first aeration channel and fell to 60 mg/l/hr or lower in the second aeration channel with the decrease in organic loading under plug flow conditions.

Depending on the number of rotors in use, variations in effluent ammonia nitrogen concentrations usually followed diurnal variations in process nitrogen loading. As indicated in Figure 6, effluent ammonia values rose to a value of 5 mg/l by 10 a.m. then decreased gradually until midnight when levels were to zero order. The variations were a function of influent ammonia concentration being high at mid-morning then tapering to values below 10 mg/l throughout the day.

Total nitrogen removals of up to 88% were realized with this process arrangement without an added carbon supplement for denitrification. Mixed liquor suspended solids ranged between 3000-7000 mg/l over a period of eight months at liquid temperatures of 12-20°C.

partially reproduced after
(Matsche & Spatzierer)

Figure 6

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(Matsche 1972)

Screen

Raw
Sewage

Degritting

Return Sludge

I

II

Clarifier
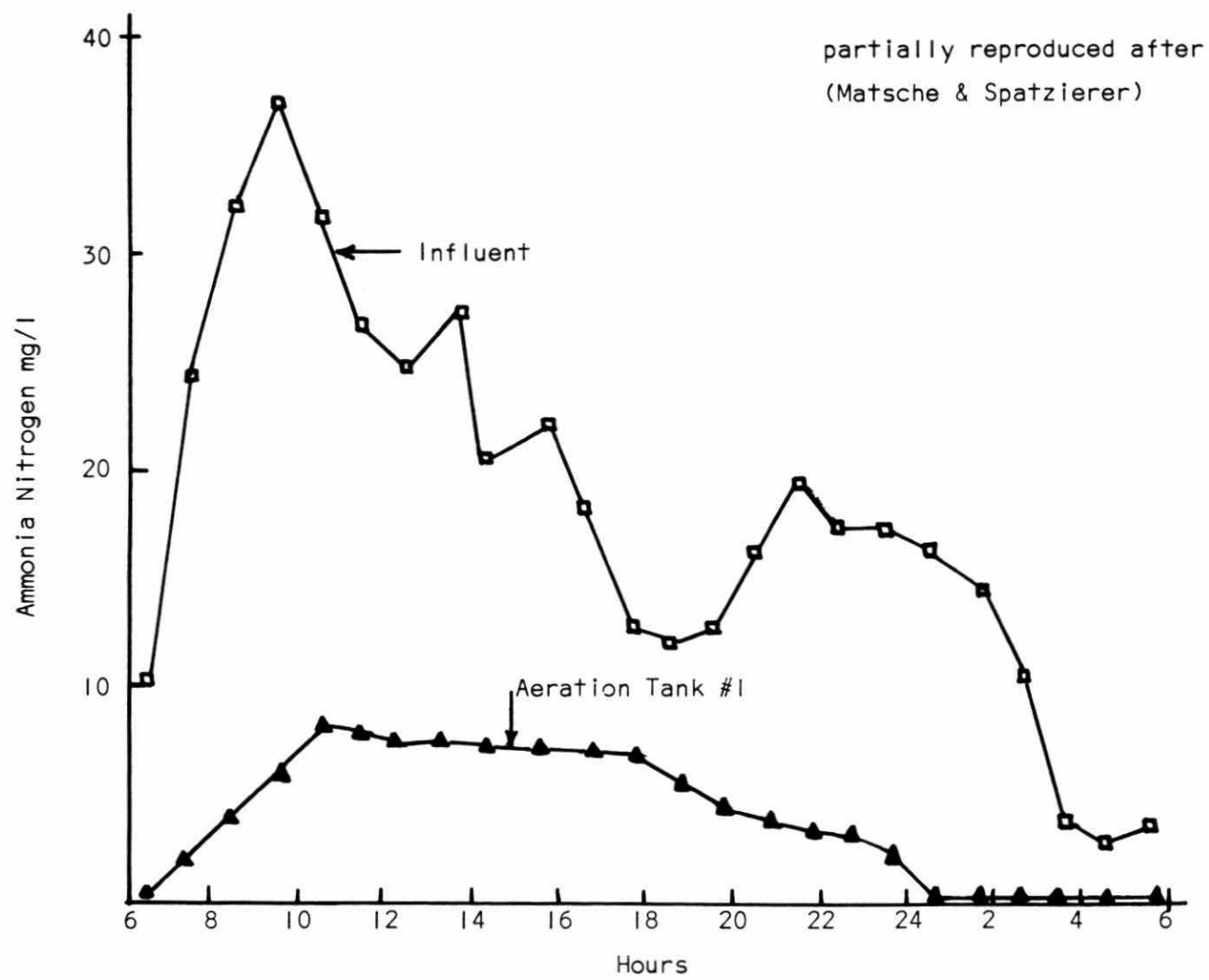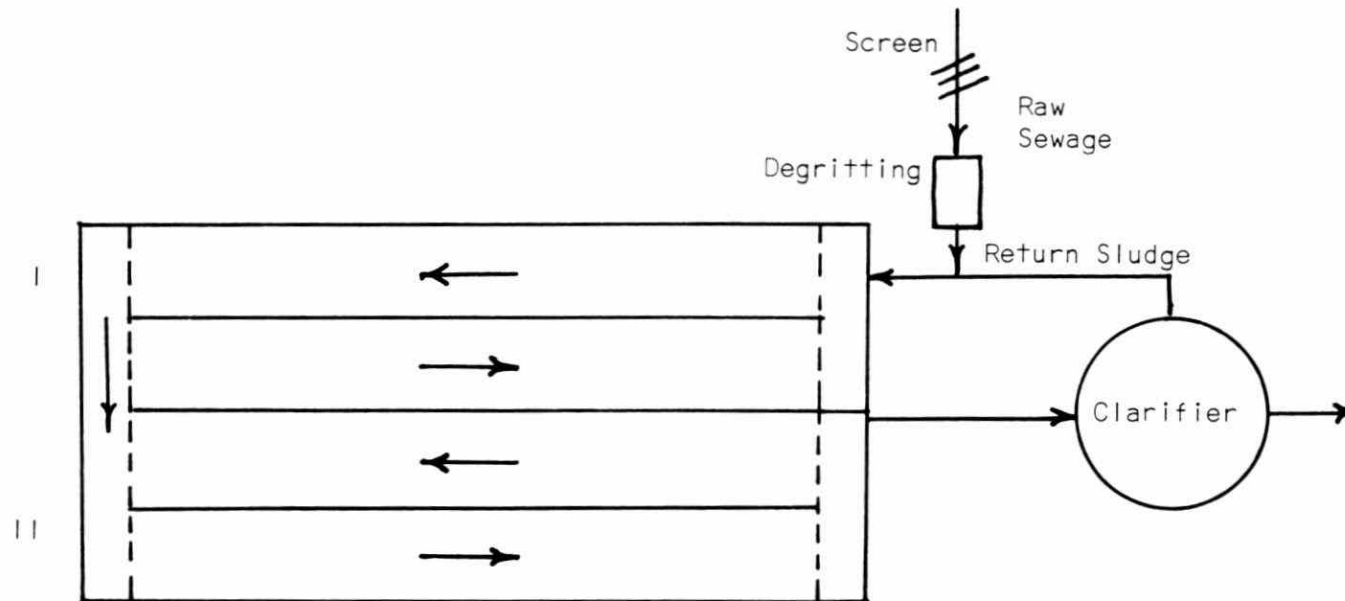
Optimum Conditions and Results

Basin Detention   - aeration 8 hr

Basin S. Solids   - 5400 mg/l

F/M ratio wt. BOD/wt. SS- 0.24

SRT   - 6-8 days

Sewage Temperature   - 18°C

Effluent Total Nitrogen - 4.0 mg/l

Effluent Ammonia   - 2.7 mg/l

Figure 5

The mean monthly organic loading ratio ranged from 0.11 to 0.24 lb BOD/lb aeration solids/day. The SRT's ranged between eight and ten days and the aeration hydraulic detention time varied from 5.5 to 8.4 hours based on mean monthly data.

EARLY STUDIES ON M.O.E. NITRIFICATION-DENITRIFICATION

Studies on biological nitrogen removal were initiated in October 1971 by the OWRC, Division of Research prior to its integration into the Ministry of the Environment. The first project entailed the use of a 900 gallon (400 litre) denitrification reactor tank into which aeration mixed liquor was fed from the Newmarket Water Pollution Control Plant (WPCP) process (6). The Newmarket plant was receiving lime addition for phosphorus removal in the primary stages of the process and information was sought as to the possible effects of lime addition on denitrification reactions. Nitrate removals of 80% were achieved in the denitrification reactor with solids concentrating to 5000 mg/l SS or methanol-to-nitrate addition ratios of 3:1. These reductions were accomplished at liquid temperatures of $8^{\circ}C$ with a reactor hydraulic detention time of seven hours. Nitrate values in the reactor effluent averaged approximately 2 mg/l under the above stated conditions.

Following the reactor study, a decision was made to move into full scale studies using the single-sludge scheme for nitrification-denitrification (7). This process design was chosen for study because of its simplicity and easy adaptation to existing biological wastewater treatment facilities. Consequently, multiple stage nitrogen removal by either biological or physical-chemical means was not chosen for investigation.

Alterations were made to the Newmarket WPCP in the summer of 1972 in which two of the existing aeration basins were connected in series for nitrification-denitrification with separate secondary clarification and sludge recycle. The one remaining aeration section with a secondary clarifier was used as a control nitrification process; both sections received primary effluent from primary clarifiers which had received lime treated raw sewage.

Full scale studies were initiated at Newmarket in September 1972 and continued until May 1973 partially financed under the Canada/ Ontario Agreement.

Good nitrogen removals were attained on occasions with methanol addition to the denitrification basin, but for the most part, both the control nitrification section and nitrogen removal section were plagued with a bulking mixed liquor. The bulking nature of the mixed liquor caused frequent losses of solids to the final effluent and made solids concentration in the denitrification basin virtually impossible. These conditions, plus the lack of facilities to achieve adequate sludge return rate, prompted the termination of the project.

It was felt at this time, that more information was required as to the control and design parameters needed to operate the single-sludge system for nitrogen removal. Consequently, a portable pilot plant was constructed for which a variety of operational choices were included in the design.

Pilot plant operations were conducted at the Brampton/Ontario Experimental Facility (O.E.F.) and the Kleinburg WPCP during the period of June, 1973 to June, 1974 (8). Many operational methods for solids control were learned during these evaluations, in particular, the technique of splitting the secondary clarifier sludge return between the aeration (nitrification) and the denitrification basins. The value of a post-aeration section between the denitrification basin and the clarifier was also realized to produce a clearer final effluent with adequate DO and free of residual methanol from the denitrification basin.

ONTARIO EXPERIMENTAL FACILITY FULL SCALE NITRIFICATION-DENITRIFICATION STUDY

History and Facility Layout

Immediately following the pilot plant experiments at Brampton and Kleinburg, a program was formulated for full scale studies using the single-stage, split-return scheme for nitrification-denitrification. By this time the Brampton Chinguacousy WPCP had been officially acquired from the South Peel Region for the purposes of research and training and the Ontario Experimental Facility (O.E.F.) was established.

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The objectives of this study were as follows:

a) To gain day to day experience in the operation of a full scale, single-stage, split-return arrangement for nitrogen removal.

b) To determine the temperature effects, if any, upon this process arrangement through seasonal variations.

c) To resolve methanol requirements of the denitrification basin for maximum nitrate-nitrogen removal.

d) To examine hydraulic detention times necessary for optimum ammonia oxidation and nitrate removal.

Alternations were made to an existing O.E.F. plant section encompassing primary clarification, mechanical aeration and final clarification, capable of handling up to four million gallons per day of sewage. A flow diagram of the plant facility is given in Figure 7.

Raw sewage is drawn from a trunk sewer which receives waste from the Town of Brampton and surrounding area. This waste can be classed as domestic-industrial with the industrial portion being derived from a variety of small to medium size companies. Sewage flows are received at variable flow to the plant within limits designated for process operation, by opening and closing a channel gate. Daily average flows can be attained through this plant section by maintaining the daytime flow at 40% above the average and allowing the nighttime flows to drop to a resultant level.

The raw sewage receives physical treatment by means of a barminutor and grit removal system and primary clarification with the raw sludge wasted to the trunk sewer. Biological treatment is performed in a seven bay mechanical aeration chamber followed by final clarification and a metered sludge return system. The final effluent and waste activated sludge are returned to the trunk sewer to be treated downline by the Lakeview WPCP in South Peel. Sewage flow monitoring is provided by a shaped capacitance probe and associated amplifiers and chart recorder. This probe is inserted in the side-well of a 6-inch parshall flume situated in the channel between the grit removal facility and the primary clarifiers.

One leg of the treatment plant was put into operation in November 1973 to test the reliability of the plant equipment for future studies and to develop a seed nitrifying sludge. The first three bays of the
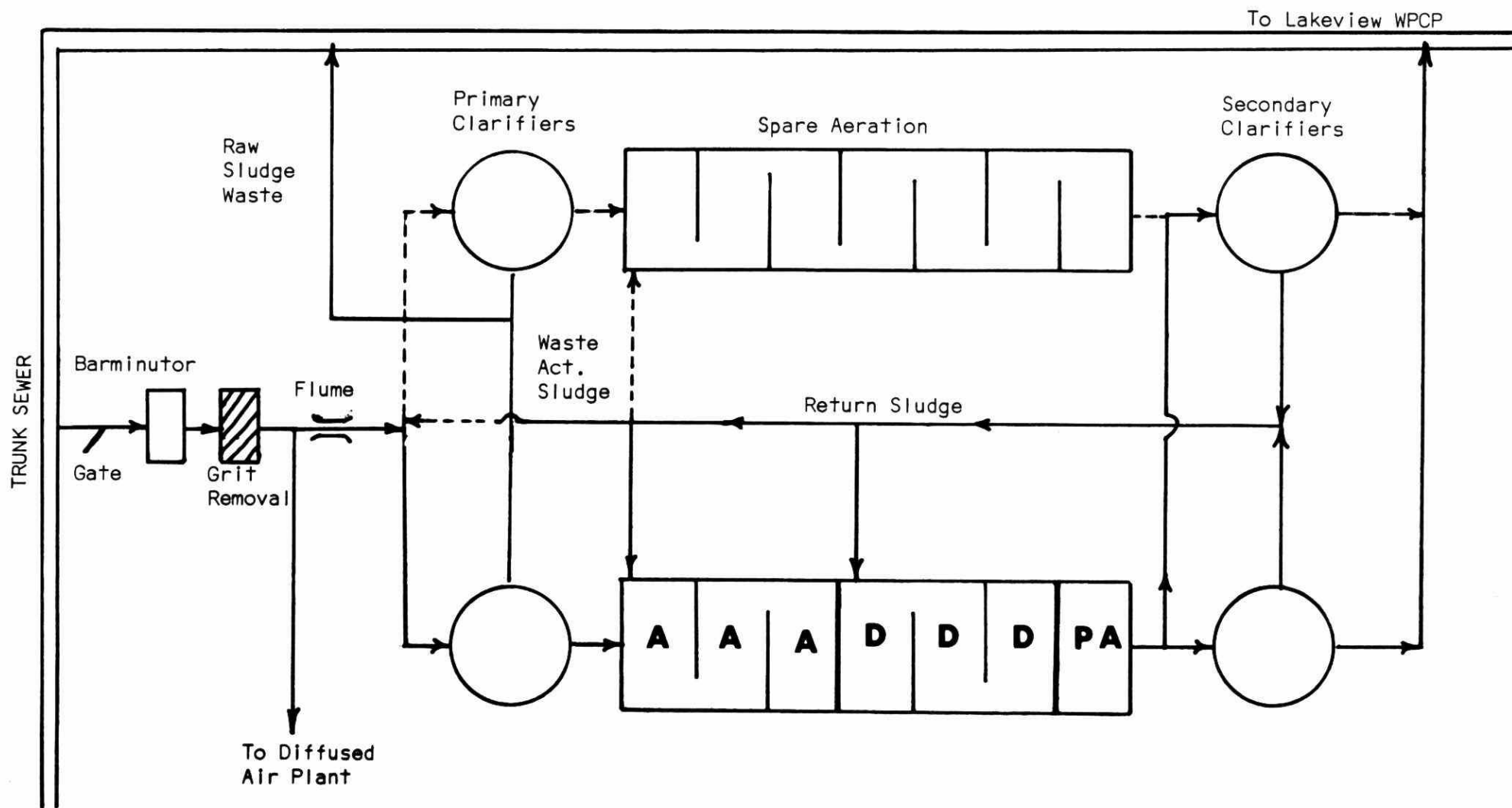
O.E.F. FLOW DIAGRAM

Figure 7

remaining leg were separated from bays 4, 5 and 6 by extending divider walls and the seventh bay was isolated via the same procedure. The aeration basin is comprised of bays 1 to 3 each having a ten horsepower mechanical aerator. Three 30 rpm mixers, formally used at Newmarket, were placed in bays 4 to 6 to facilitate slow mix denitrification basins and the seventh bay served as a post-aeration chamber prior to final clarification utilizing one ten horsepower mechanical aeration unit. This physical arrangement was used for the first four programs of this study but alterations were made to change basin volumes in subsequent programs.

Design of Study

The O.E.F. project has been classified into various programs, each entailing a minimum of six weeks process operation. The programs completed or in progress are outlined as follows:

Program 1   - Basin hydraulic det. - aeration 8 hr

- denitrification 8 hr

- post-aeration 2 hr

Mean liquid temperature of $18^{\circ}C$

Program 2   - Basin hydraulic det. - aeration 6 hr

- denitrification 6 hr

- post-aeration 1.2 hr

Mean liquid temperature of $16^{\circ}C$

Program 3   - same as program 2 only at mean liquid temperature of $10^{\circ}C$

Program 4(a) - same as program 1 only at mean liquid temperature of $10^{\circ}C$

4(b) - same as program 4(a) only hydrogen peroxide was added to the process in an attempt to control sludge bulking

Program 5(a) - Basin hydraulic det. - aeration 8 hr

- denitrification 2 hr

- post-aeration 4 hr

Mean liquid temperature of 20°C

5(b) - Basin hydraulic det. - aeration 10 hr

- denitrification 2.5 hr

- post-aeration 4.8 hr

Mean liquid temperature of 22°C

5(c) - Basin hydraulic det. - aeration 10.5 hr

- denitrification 2.6 hr

- post-aeration 5.2 hr

No methanol was added for denitrification during this sequence.

Mean liquid temperature of 18°C

Program 6   - Studies in progress.

Basin hydraulic det. - aeration 10 hr

- denitrification 4 hr

- post-aeration 2 hr

Ferric chloride added to the post-aeration basin at a dosage range of 10-15 mg/l as Fe.

Constant Process Parameters

The overall process parameters applied to programs 1 to 5 are as follows:

(a) Aeration F:M ratio maintained at 0.15 to 0.3 lb BOD/lb VSS.
(b) Aeration ML dissolved oxygen between 2-4 mg/l.
(c) Secondary clarifier sludge return to basins was 100 to 200% of plant flow and split between aeration and denitrification processes.

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(d) ML suspended solids concentration in the denitrification basin was maintained between 4500-6000 mg/l.

(e) Methanol was added in an amount required to maintain less than 1 mg/l of NO_3-N in the denitrification basin.

PROJECT OPERATIONS (PROGRAMS 1 TO 4)

The nitrification-denitrification process was initiated in June, 1973 by transferring established aeration mixed liquor from the spare aeration section to the reconstructed nitrogen removal facility. After approximately three weeks of observation, it became evident that there was considerable backflow of denitrification ML to the aeration basin via the flow patterns produced by the draft-tube type mechanical aerators. A deterioration in nitrification reactions resulted with accompanying increased oxygen utilization rate. Consequently, the process was reverted to the spare aeration section while revisions were made to the basin wall divisions. The walls between the aeration and the denitrification basin, and the denitrification and the post-aeration basin were sealed with the exception of a two-foot square opening and a flap-gate.

The nitrogen removal process was restarted in July similar to the aforementioned procedure and the facility operated well without noticeable inter-basin back-mixing, throughout the various programs.

ROUTINE SAMPLING AND ANALYSES

(a) Twenty-four hour composite samples were taken five days per week of raw sewage and primary and secondary clarifier effluents. Portions of these composites were filtered for soluble component analyses.

(b) Filtered composite grab samples were made of the aeration and denitrification basins ML daily for dissolved chemical analyses.

(c) Samples were taken to the MOE Central Laboratory Facility each day for a comprehensive series of chemical analyses. These analyses were as follows:

i) Unfiltered - biochemical oxygen demand (BOD), chemical oxygen demand, total phosphorus, alkalinity, total Kjeldahl nitrogen and suspended solids.

TABLE I

PLANT VOLUMES AND DETENTION TIMES

Volumes

Primary	–	125,000 IG
Final	–	175,000 IG
TOTAL REACTION CHAMBER	–	500,000 IG

Programs I to 4

Nitrification Vol.	–	214,000 IG
Denitrification Vol.	–	214,000 IG
Post-Aeration Vol.	–	72,000 IG

Mean Clarifier Upflow Rates GPD/ft^2

PI & 4 Primary	–	346
Secondary	–	240
P2 & 3 Primary	–	438
Secondary	–	304

MEAN DETENTION TIMES AND FLOWS REQUIRED

Nitrification and Denitrification

(a) 8 hr 28,333 G/hr or 0.68 Imgd

(b) 6 hr 35,700 G/hr or 0.86 Imgd

ii) Filtered - inorganic and organic carbon, phosphorus and
ammonia, nitrite and nitrate nitrogen.

d) On-site tests needed for process control were performed
at the O.E.F. laboratory each day as follows:

Mixed liquor 1/2 settling tests; suspended and volatile
solids, oxygen utilization rates and pH.
Mixed liquor filtrates, ammonia and nitrate nitrogen.

RESULTS (PROGRAMS 1 AND 2)

Nitrification

The ammonia conversion (nitrification) in the aeration basin
averaged 75% during Program 1 with a further increase in ammonia removal
taking place in the latter stages of the process. An average F:M ratio
of 0.16 lb primary BOD/day/lb aeration VSS was produced and dissolved
oxygen concentrations across the aeration basin were maintained at 2 to
5 mg/l. Residual concentrations of ammonia nitrogen escaping the aeration
chamber were usually nitrified in the denitrification basin or the post-
aeration bay prior to secondary clarification. Results from a sampling
and test survey on the process revealed that aeration ammonia breakthrough
occurred from 11 a.m. to midnight when the sewage flows were normally
40% above the average daily level.

Aeration nitrification through Program 2 was surprisingly improved
to an average of 88% ammonia conversion. The reasons for this improvement
are thought to be a possible increase in nitrification bacteria in the
process sludges, or better control over peak sewage loading periods.
Moreover, the unaccountable loss of soluble nitrogen in the aeration basin
was 49% of the primary effluent as compared to 21% in Program 1. A more
in-depth analysis of the project data should reveal some clues as to the
mechanisms of the increased nitrogen loss. As in Program 1, residual
ammonia escaping the aeration basin was converted and/or lost in the
remaining process stages. Sludge growth rate increased during Program
2 and more frequent sludge wasting was required to maintain solids control
in both the aeration and denitrification chambers. The solids retention
times (SRT) were 41 and 11 days in Programs 1 and 2, respectively.

Denitrification

Soluble nitrogen removal (ammonia, nitrite and nitrate) proceeded at a high rate throughout both programs with methanol added to the denitrification chamber as a supplementary carbon source. A methanol-to-aeration nitrate nitrogen addition ratio of 1 to 2:1 was maintained in the first denitrification bay to keep the nitrate content in the denitrification chamber below 1 mg/l. Sampling surveys along the breadth of the denitrification chamber showed that the majority of nitrite and nitrate reduction appeared in the first bay.

Soluble N values of the final effluent closely approximated those of the denitrification chamber, although occasionally nitrification or denitrification of residual nitrogen did occur in the secondary clarifier.

The pounds of methanol required per lb of nitrate in the denitrification chamber for Programs 1 and 2 were 0.9 and 1.2, respectively, with the methanol requirements based on total soluble nitrogen removed being of a similar relationship. A summary of results for Programs 1 and 2 is given in Table 2.

Microbiological Observation and Sludge Conditions

A variety of protozoa were observed to populate both the aerobic and semi-anaerobic sludges (*aspidicus vorticella* and *opercularia*) during these programs. As the programs progressed *opercularia* became the most dominant protozoa and three varieties of filamentous bacteria, *sphareotilus*, *actinomycetes* and *vitreoscilla*, increased in numbers. The quantity of filamentous forms in the process sludge remained at a medium level compared to past observations of filamentous sludges. The sludge volume index (SVI) of both the aeration and denitrification mixed liquors were consistently between 200-300 throughout both programs. It is felt that the combination of filamentous bacteria content and changes in sludge characteristics due to anaerobic sludge detention created a poor settleable floc. These sludge characteristics, along with raised hydraulic loadings, produced bulking conditions with a resultant loss of solids to the final effluent at clarifier loadings of more than 450 gpd/ft^2. Increases of oxygen utilization rate were observed in the denitrification mixed liquor

TABLE 2

O.E.F. FULL-SCALE NITRIFICATION-DENITRIFICATION STUDY

Summary of Results (Average Values)

Measurement	Program 1 8-Hr Detention	Program 2 6-Hr Detention
Raw Sewage		
Flow 1mgpd	0.687	0.825
BOD	166	206
SS	223	346
Tot. Kj	38	40
Tot. P	8.2	9.2
Primary Effluent		
BOD	89	99
NH_3-N	16	21
Sol. TC	99	92
Alk.	216	230
Aeration Chamber		
NH_3-N	3.8	2.5
NO_2-N	1.0	0.4
NO_3-N	7.9	7.4
SS	2311	2268
VSS	1841	1592
Resp. Rate mg/l/hr	19	36
Denitrification Chamber		
NH_3-N	0.8	0.8
NO_2-N	0.25	0.27
NO_3-N	1.0	1.7
SS	4241	4647
VSS	3594	3386
Resp. Rate mg/l/hr	45	52
Average Liquid Temp. ^{o}C	16	18

Table 2 - cont'd

Measurement	Program 1 8-Hr Detention	Program 2 6-Hr Detention
Final Effluent		
BOD	7.0	9.6
SS	18	20
Tot. Kj	3.1	2.8
Tot. P	3.8	4.6
NH_3-N	0.6	0.6
NO_2-N	0.18	0.05
NO_3-N	1.3	1.2
Sol. TC	53	50
Alk.	160	138
F:M ratio	0.16	0.24
Methanol Addition lb/day	48	75

under anoxic conditions. These increases were 30 to 60% above aeration basin levels for programs 1 and 2, respectively.

RESULTS (PROGRAMS 3 AND 4)

Nitrification

Utilizing the six hour detention mode at mean sewage temperatures of $10^{\circ}C$ in program 3 did not deter ammonia conversion in the aeration basin as 87% efficiency was maintained. The reason for this may be that a more constant and slightly lower F:M ratio was being maintained, plus that the aeration mixed liquor dissolved oxygen levels were more uniform along the length of the aeration basin (Table 3).

A marked improvement was shown in nitrification efficiency in program 4 with the increase of aeration detention time to eight hours. An unaccountable soluble nitrogen loss of 40% experienced in program 3 was reduced to 23% in program 4 with augmented nitrate formation.

STR's were difficult to calculate with any accuracy in program 3 because of sludge losses in the secondary effluent due to bulking on weekends caused by overnight stormwater infiltration. Flows were more consistently controlled during program 4 and a SRT of 30 days was computed.

Denitrification

Final effluent nitrate-nitrogen levels were maintained at about 1 mg/l throughout programs 3 and 4 at lower sewage temperatures and reduced methanol addition. This increase in process nitrogen removal can be attributed to higher MLSS concentrations in the denitrification basin, particularly in program 4, lowering the methanol requirements for nitrate reduction. The resultant methanol requirements for programs 3 and 4 were consequently lower at respective values of 0.9 and 0.4 lb methanol to lb nitrate-nitrogen. Heightened oxygen utilization rates were not experienced during these programs as compared to programs 1 and 2, possibly due to lower F/M conditons and liquid temperatures.

HYDROGEN PEROXIDE ADDITION TO O.E.F. PROCESS

Although excellent nitrogen, BOD and COD removals were obtained throughout the aforementioned programs, a persistent bulky mixed liquor

TABLE 3

O.E.F. FULL-SCALE NITRIFICATION-DENITRIFICATION STUDY

Summary of Results (Average Values)

Measurement	Program 3 8-Hr Detention	Program 4 6-Hr Detention
Raw Sewage		
Flow lmgd	0.822	0.692
BOD	162	161
SS	238.8	256
Tot. Kj	31	29
Tot. P	7.1	6.9
Primary Effluent		
BOD	85	73
NH_3-N	18	13
Sol. TC	99	96
Alk.	226.5	235
Aeration Chamber		
NH_3-N	2.3	1.0
NO_2-N	0.30	0.04
NO_3-N	8.5	10.3
SS	2201	1818
VSS	1768	1455
Resp. Rate mg/l/hr	21.4	15.4
Denitrification Chamber		
NH_3-N	0.74	0.22
NO_2-N	0.07	0.07
NO_3-N	0.97	0.79
SS	4723	5551
VSS	3784	4477
Resp. Rate mg/l/hr	26.3	23.1
Avg. liquid Temp. $^{\circ}C$	10.1	10.3

Table 3 - cont'd

Measurement	Program 3 8-Hr Detention	Program 4 6-Hr Detention
Final Effluent		
BOD	7.3	3.5
SS	10	8.3
Tot. Kj	2.31	1.53
Tot. P	3.49	3.23
NH_3-N	.86	.11
NO_2-N	.06	.04
NO_3-N	1.1	1.2
Sol. TC	50.1	53.2
Alk.	134	181
F:M ratio	0.19	0.16
Methanol additon lb/day	63	31

was experienced in all the process basins. A four hour detention mode
had been considered by increasing the sewage flow to the process, but
secondary clarifier overflow rates at this time could not exceed a mean
daily value of 300 gal/ft^2 without high losses of solids from the process.
The quantities of filamentous organisms were high and deflocculation of
the mixed liquor occurred under anoxic conditions; the combination of
these factors produced aeration SVI's of up to 500 mg/l.

During the month of May, 1975, hydrogen peroxide was added to
the O.E.F. nitrogen removal process at dosages ranging from 25 to
100 mg/l. Some improvement was shown as aeration SVI's dropped to values
of approximately 250, but a quick recovery of high SVI was experienced
with cessation of peroxide addition.

During and following peroxide addition periods of one to five
days, a rapid depreciation in nitrification and subsequently denitrification
reactions occurred (Figure 8). A more in-depth discussion of this study
is described in a separate report by the author (9).

PROGRAMS 5 a, b, c and 6

Following the peroxide addition experiment, a process recovery
period of four weeks was necessary to regain full nitrification in the
aeration basin. This was partly accomplished by switching the process
mixed liquor to the spare aeration basin for conventional operation.
Physical alterations were made to the nitrogen removal facility to increase
the volume of the aeration and post-aeration basins and decrease
the volume for denitrification. The purposes of these changes were to
stimulate process mixed liquor flocculation by increasing the period of
aerobic activity prior to, and following the anoxic sequence, to evaluate
the methanol requirements of the reduced denitrification volume and
determine the nitrogen removal efficiency of this arrangement without
methanol supplementation. Step feeding of primary effluent to the
first half of the aeration basin was also introduced to equalize the
loading to the basin.

Program 5 (a)

A five week study initiated in late June was carried out at
sewage flows averaging 0.833 Imgpd and liquid temperatures of 20°C.

Figure 8

O.E.F. H$_2$O$_2$ Study

P.A. – pre-addition period
A$_1$ – 25-98 mg/l H$_2$O$_2$
R$_1$ – non-addition period
A$_2$ – 98 mg/l H$_2$O$_2$
R$_2$ – non-additon recovery

Nitrogen Removal (%)

% Total Nitrogen

% Soluble Nitrogen

April

May

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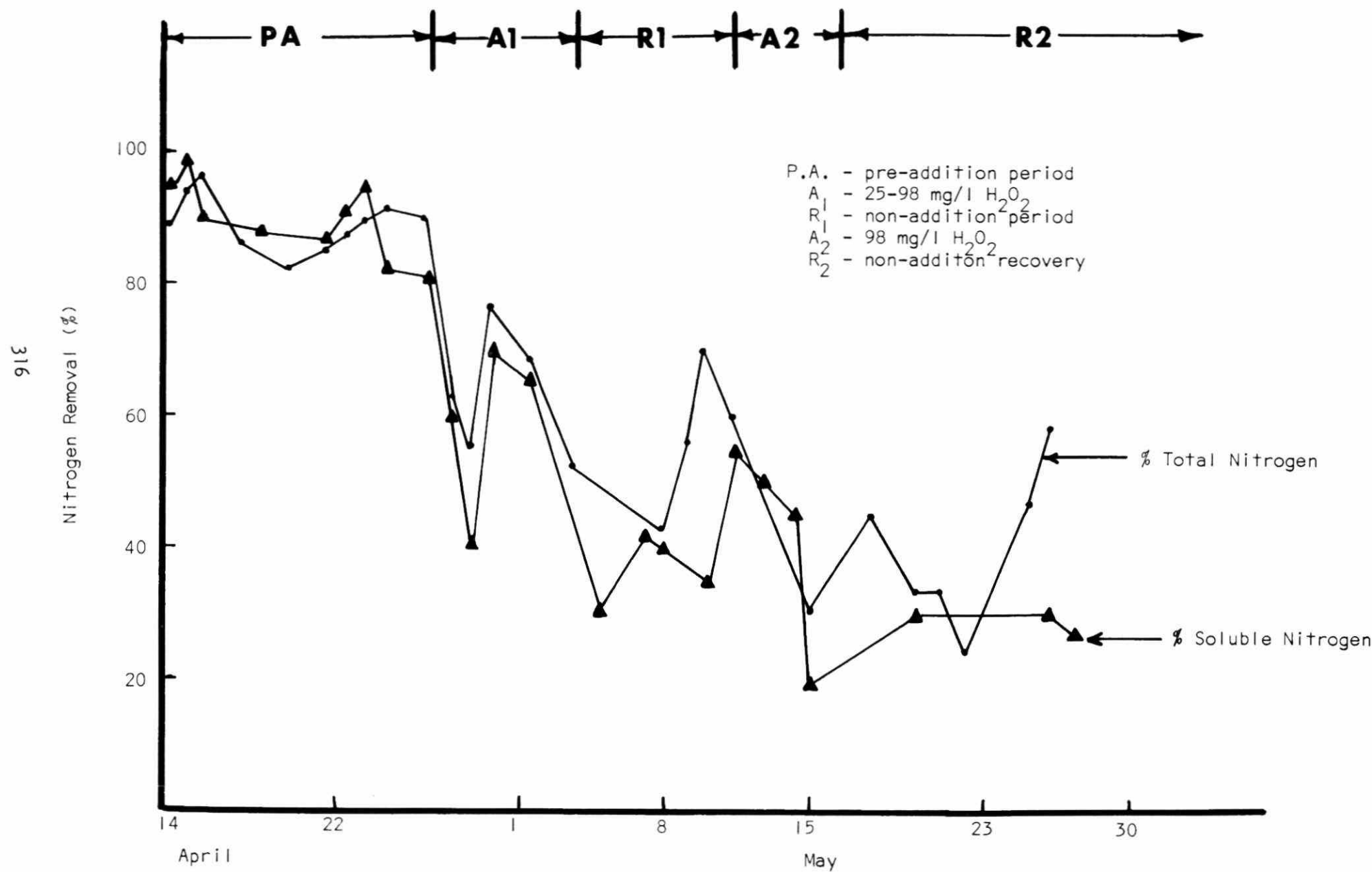
Through this phase full ammonia conversion was accomplished at a mean
F/M ratio of 0.18 lb BOD/lb VSS with a basin hydraulic detention time
of 8.2 hours.

Denitrification reactions did deteriorate and nitrate levels
in the denitrification basin approached 3.0 mg/l even with a methanol-to-
nitrate addition ratio of 3 to 1, and a hydraulic detention of two hours.
Secondary effluent nitrate results showed a mean of 1 mg/l; consequently
a further nitrate reduction occurred either in the post-aeration or in
the secondary clarifier sludge blanket.

Program 5 (b)

During the month of August, the sewage flows were lowered to
0.7 Imgpd to give hydraulic detentions in the aeration and denitrifica-
tion basins of 10 and 2.5 hours, respectively. Nitrification reactions
in the aeration basin dropped somewhat giving ML residual ammonia values
approaching 2 mg/l and nitrate levels less than 8 mg/l.

Adequate nitrate removal was also difficult to achieve as a
mean methanol-to-nitrate addition ratio of 6 to 1 did not reduce denitri-
fication ML values below 2 mg/l. Furthermore, an extra 1 mg/l of nitrate
was added to the secondary effluent via oxidation of process residual
ammonia in the post-aeration basin.

Program 5 (c)

The third phase of program 5 was run without methanol addition
to the denitrification basin at a mean sewage temperature of 18°C and
similar flows and basin detention to P 5 (b). An immediate improvement
in ammonia conversion took place in the aeration basin with the cessation
of methanol application to the process. This suggests that a toxicity
was being imposed upon the nitrification bacteria during the previous
phase with the large amounts of methanol addition.

Conversely, denitrification was affected to a point where up to
5 mg/l nitrate-nitrogen was realized in the denitrification basin mixed
liquor. Denitrification resulted to some degree in the secondary
clarifier but effluent values did not fall much below 4 mg/l nitrate
nitrogen.

A summary of results for program 5 is given in Table 4.

Program 6

For program 6, which is now being studied, the denitrification basin was increased to four hours and the post-aeration reduced to two hours detention, respectively. Ferric chloride addition has been initiated to the post-aeration basin in an attempt to improve clarifier sedimentation and obtain phosphorus removal. Early results show effluent total phosphorus levels approximating 1 mg/l with 12 mg/l ferric chloride as Fe added to the process. No deterioration in nitrification or denitrification reactions has been observed and nitrogen reductions with minimal methanol supplementation are proceeding at efficiencies obtained in programs 1 to 4. A decrease in SVI is gradually being realized in the aeration mixed liquor.

SCREEN CENTRIFUGE TRIAL

The changes in facility design instituted in program 5 failed to make any significant difference in process mixed liquor flocculation. Recent studies by Ferbach and Tchobanoglous have shown that changes in floc characteristics plus a disturbance in interfloc filamentous organism network, can be achieved by using a screen centrifuge concentrator on activated sludge (3). Moreover, a 50% reduction in solids can be achieved by this unit and if it is installed between the aeration basin outfall and the secondary clarifier a considerable reduction in clarifier solids loading will be achieved.

During program 5 (c) a mobile screen centrifuge* capable of handling 100 gal/min of sewage was operated in parallel to the nitrogen removal facility. The unit was operated eight hours/day on denitrification mixed liquor and a comparison was made between influent and effluent mixed liquor characteristics.

The results of this evaluation were quite encouraging as dramatic solids reduction and reflocculation occurred on processing the denitrification mixed liquor. Microscopic examination of mixed liquors confirmed the effects on filamentous organisms as experienced by Fernbach and Tchobanoglous and the extensive filamentous networks were disturbed and

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* The screen centrifuge was made available with a member of staff by Sweco Canada Ltd.

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TABLE 4

PROGRAM 5 - SUMMARY OF RESULTS

MEAN SOLUBLE NITROGEN MG/L

Program Phase	Primary Effluent NH_3	Aeration Basin			Denitrification Basin			Secondary Clarifier		
		NH_3	NO_2	NO_3	NH_3	NO_2	NO_3	NH_3	NO_2	NO_3
P5 (a) Meth./NO_2 3:1	19	0.89	0.79	9.9	0.56	0.12	2.8	0.34	0.06	1.0
P5 (b) Meth./NO_3 6:1	22	1.6	0.05	7.9	1.3	0.02	2.1	1.5	0.1	2.7
P5 (c)	18	0.18	0.02	11.1	0.11	0.07	4.5	1.4	0.15	3.9

Note: All nitrogen values reported as N

the organisms seemed to be broken and scattered away from the floc particles.

SURVEY OF MINISTRY OF THE ENVIRONMENT WPCP's

To complement ongoing studies on nitrification-dentrification, a data survey of existing wastewater treatment facilities was undertaken to determine the efficiency of ammonia conversion presently being accomplished. The distribution as presented in Table 5 was calculated from the 1973 MOE WPCP operations data.

TABLE 5

MINISTRY OF THE ENVIRONMENT WPCP's 1973

DISTRIBUTION OF NITRIFICATION EFFICIENCY

(Results mg/l N)

Effluent NH_3 Range	Effluent NO_3 Range	% of Plants
	Extended Aeration	
0 - 3	7 - 10	44
4 - 10	1 - 6	34
11 - 20	less than 1	22
		100
	Conventional	
0 - 3	9 - 12	41
4 - 10	4 - 8	30
11 - 20	0 - 3	29
		100

The data suggest that considerable nitrification is being attained in Ministry plants, in particular in extended aeration facilities, and that possible nitrate nitrogen removal may be taking place in at least 20% of the extended aeration plants. The quantity of data available for conventional plants far exceeded that for extended plants and consequently an accurate nitrification comparison cannot be made between these process

schemes. Plants using the extended aeration mode have generally been observed to have low ammonia effluents.

DISCUSSION

Evaluation of the full scale, single-sludge, nitrification-denitrification processes shows that total nitrogen reductions of 90% can be accomplished if adequate hydraulic detention time and/or adequate solids concentration can be maintained in the aeration (nitrification) and denitrification basins. These reductions generally would produce secondary effluents containing total nitrogen of less than 5 mg/l with the organic nitrogen portion being 2-3 mg/l under ideal conditions. Some deterioration will probably be seen in nitrate removal when sewage temperatures drop below 15°C if denitrification solids are not maintained above 4000 mg/l SS and/or methanol addition is not implemented. Most of the processes described in this report offered SRT's exceeding 6 and often approaching 30 days; consequently, an active mass of nitrogen bacteria was probably retained within each process system.

Present data analyses of the O.E.F. study do not indicate a deterioration in nitrification efficiency on moving from the eight hour to the six hour basin detention mode either at 20 or 10°C mean sewage temperature. Nitrification was affected slightly when methanol-to-nitrate addition ratios exceeded 3:1.

The luxury uptake of phosphorus was not observed as in Barnard's studies (1) with a similar process; only a phosphorus release was noted as soluble phosphorus levels usually rose 1 to 2 mg/l in the denitrification stage.

The bulking sludge constantly experienced in the O.E.F. process could pose problems with clarification and increased capacity may be required. Lowered SVI's may be achieved by the addition of ferric chloride for phosphorus removal presently underway through the formation of a higher density chemical-biological sludge. This in turn could improve secondary clarifier sedimentation and allow higher overflow rates than the maximum peak of 400 gal/ft^2/day attained for the filamentous sludge. Past experiences with these clarifiers, even on a normal SVI sludge, indicate that the surface area to depth ratio (60 ft diameter/ 10 feet deep) is too high and under these conditons the sludge blanket

can be easily disturbed by accelerating flow conditions or surface wind action.

A viable alternative to changes in clarifier design may be the incorporation of a screen centrifuge to replace the post-aeration section. Pilot experiments showed a marked improvement in denitrification mixed liquor settleability along with raised DO levels using this device. Consequently, a full scale installation of this unit could improve secondary clarifier performance without chemical addition.

The secondary clarifier effluent remained exceptionally clear and relatively free of colloidal material throughout all programs, with effluent BOD and SS frequently below 10 mg/l as long as the overflow rate was not allowed to exceed 400 gal/ft^2/day. Furthermore, soluble organic carbon analyses performed routinely on secondary clarifier effluent did not indicate methanol carryover from the denitrification basin.

A complete process cost analyses was not attempted in this report but out of five programs of study at the O.E.F., six hour aeration, four hour denitrification and two hour post-aeration detentions appear to be adequate. Secondary clarification overflow rates of 300 gal/ft^2/day based on a mean daily flow would be required due to the inherent bulky nature of the process mixed liquor, if some way to improve sedimentation is not found. A possible alternative process design would be to replace the post-aeration basin with a screen centrifuge concentrator and recycle the concentrated sludge to the denitrification basin. A proposal for a full scale installation of this type is presently being considered by the Wastewater Treatment Section for 1976. Cost implications of such an alternative would thereby be evaluated.

A low sludge production was experienced with the O.E.F. nitrogen removal process, as sludge wasting was required only once or twice per month. This, therefore, could result in a cost saving for sludge treatment. Generally, the split-return, single-sludge process was simple to operate on a day to day basis and the man-hours required would not be more than a conventional activated sludge plant. Nitrate nitrogen levels could be easily controlled in the denitrification basin by performing daily nitrate analyses using a field test kit and the methanol feed rate adjusted to maintain maximum denitrification. Methanol require-

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ments based on the current price of 0.66¢/gallon would amount to 4 and 7¢/1000 gallons of sewage treated with eight and six hour detention modes, respectively.

CONCLUSIONS

1. The single-sludge scheme for nitrification-denitrification offers an efficient and economical way of expanding existing waste treatment facilities for nitrogen removal.

2. The single-sludge, split-return process offers consistent effluent total nitrogen below 4 mg/l with minimal methanol addition and operator attention under varying climatic conditions.

3. At low temperatures, methanol requirements can be reduced by increasing the denitrification detention time, solids concentration and resultant process SRT.

4. Chemical or physical treatment may be required to produce a denser sludge and/or to reflocculate the denitrification mixed liquor prior to secondary sedimentation.

5. A process approaching the extended aeration mode of operation offers the opportunity of simultaneous nitrification and denitrification.

6. Approximately 60% of existing M.O.E. WPCP's require some form of facility alteration to achieve effluent ammonia levels below 3 mg/l.

REFERENCES

1. Barnard, J.L., "Cut P and N without Chemicals", Water and Wastes Engrg., Part 1, 33 (July 1974), Part II, 41 (Aug. 1974).

2. Bishop, D.F., Heidman, A., and Stanberg, J.B., "Single-Stage Nitrification-Denitrification", Paper presented at the 47th Annual Conf. of the Water Pol. Control Fed., Oct. 6-11, 1974, Denver, Colorado.

3. Fernbach, E. and Tchobanoglous, G., "Centrifuge Screen Concentrator for Activated Sludge Process", Water and Sewage Works, Jan. and Feb., 1975.

4. Matsche, N.F., "The Elimination of Nitrogen in the Treatment Plant of Vienna-Blumental", Water Research, Pergamon Press, Vol. 6, p. 485-486, 1972.

5. Nicholls, H.A., "Modification of Extended Aeration Plants in Johannesburg, South Africa to Achieve Denitrification", IAWPR Spec. Conf. on Nitrogen as a Water Pollutant, Aug. 1975, Denmark.

6. Smith, A.G., "Denitrification Reactor Studies in a Lime Treated Sewage Plant", M.O.E. Report No. W2029, July, 1972.

7. Smith A.G., "Full-Scale Nitrification-Denitrification Studies at the Newmarket Nutrient Removal Facility", Unpublished report under Canada-Ontario Agreement Project No. 71-1-20.

8. Smith, A.G., "Nitrification-Denitrification Pilot Plant Studies at Brampton and Kleinburg", Unpublished report under Canada-Ontario Agreement Project No. 71-1-20.

9. Smith, A.G., "Hydrogen Peroxide Addition to the Ontario Experimental Facility Nitrification-Denitrification Process", Pre-publication, Ministry of the Environment, Pollution Control Branch.

10. Wuhrmann, K., "Objectives, Technology and Results of Nitrogen and Phosphorus Removal Process", Advances in Water Quality Improvement by G. Loyda and Eckenfelder, Vol I, Pg. 21-48, 1968.

11. Wuhrmann, K., "Effect of Oxygen Tension on Biochemical Reactions in Sewage Purification Plants", Proc. 3rd Conf. Biol. Waste Treatment, 1960, Advances in Biological Waste Treatment, Oxford Pergamon Press, 1963.

THE USE OF ACTIVATED CARBON IN
CONVENTIONAL ACTIVATED SLUDGE PROCESSES

by

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INTRODUCTION

The use of activated carbon for the adsorptive removal of soluble organic material from wastewaters is being investigated in several parts of the world. This paper is a brief summary of the continuing work with activated carbon being undertaken by the Research Branch of the Ontario Ministry of the Environment and the Environmental Protection Service of Environment Canada.

Powdered activated carbon (PAC) was applied in a full size conventional activated sludge process prior to both primary and secondary clarification. PAC was also used in physical-chemical treatment processes, full size and pilot scale, in conjunction with chemical coagulation and clarification. Granular activated carbon (GAC) was applied to a secondary (biological) effluent and to effluents from chemical clarification processes using pressurized contact columns which ranged in size from five inches in diameter to five feet in diameter. Some experiments are complete, others are still in progress. In addition, experimentation on carbon regeneration is presently being planned and initiated.

Figure 1 illustrates the potential applications of activated carbon, in powdered and granular form, to the activated sludge process. Granular activated carbon may be used efficiently as an adsorbent of residual or refractory soluble organic materials which escape the secondary treatment process. A filter unit preceding the adsorption

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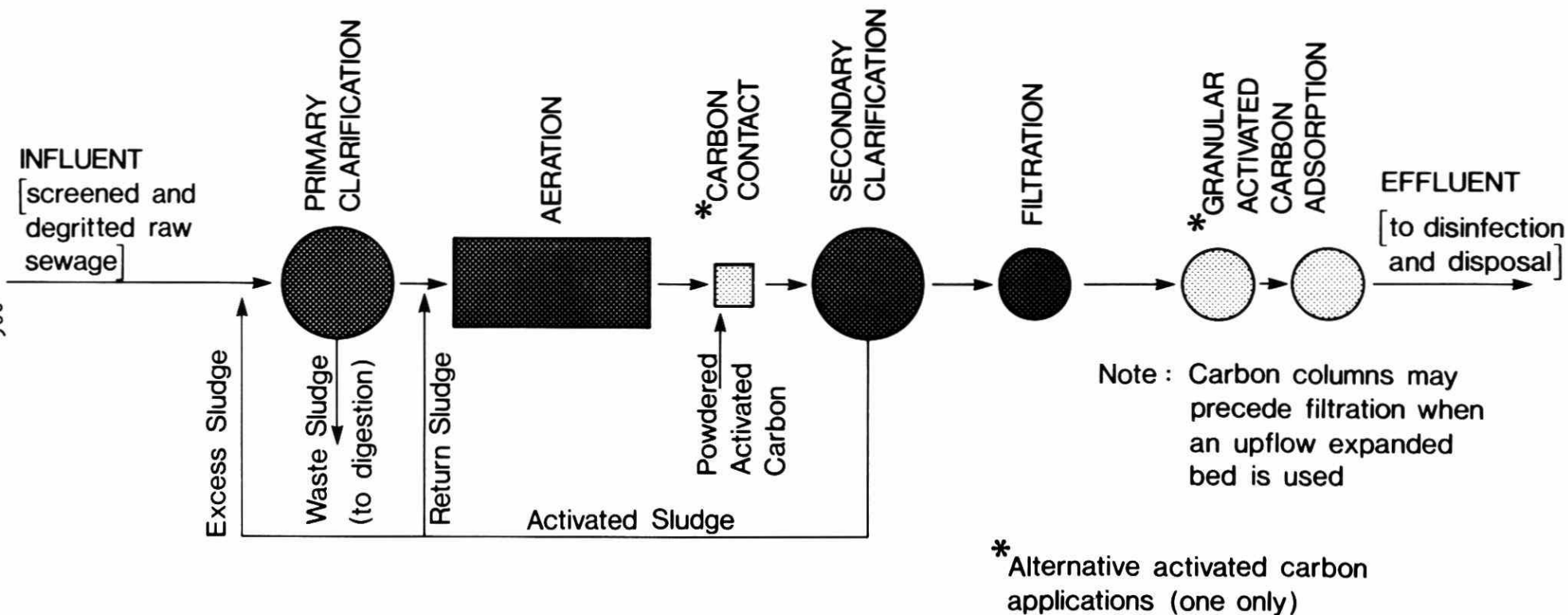
FIGURE 1

ACTIVATED CARBON APPLICATIONS TO THE ACTIVATED SLUDGE PROCESS

unit is illustrated in Figure 1; alternative schemes may combine filtration with downflow carbon adsorption or use filtration as a final polishing step following carbon adsorption. Powdered activated carbon may be applied to the wastewater stream prior to secondary clarification. This latter application suffers from problems associated with solids carry-over in the clarifier effluent and activated carbon reclamation from the waste sludge.

Figure 2 illustrates the potential applications of activated carbon in physical-chemical treatment processes. Granular carbon may be used following the removal of most of the suspended solids. The filtration stage may not be required before carbon contact depending on the efficiency of upstream units. Powdered activated carbon may be applied to the raw sewage or in separate contact units following primary clarification.

POWDERED ACTIVATED CARBON

PAC in Conventional Activated Sludge Processes (Bolton WPCP)

The Ontario Ministry of the Environment has, since 1973, been studying the use of activated carbon for soluble organic removal in a 0.5 MGD conventional activated sludge plant at Bolton. The Bolton Water Pollution Control Plant consists of two parallel 0.25 MGD plants and is thus easily adaptable for the performance of controlled experiments. The final clarifiers are rectangular [10.9 x 3.6 x 3.6 m (36 x 12 x 12 feet)] and have a 3.1 hour detention time and surface loading rate of 1.0 m^3/m^2h (580 gpd/ft^2) at design flow. Powdered activated carbon (Aqua Nuchar A) in conjunction with a coagulant, was applied to one half of the plant at the downstream end of the aeration basin prior to final clarification. The control half of the plant was operated under the same loading conditions, with the same coagulant dose but without activated carbon.

Process performance results at carbon doses of 25 and 50 mg/l and an alum dose of 150 mg/l are presented in Tables 1 and 2. These results indicate that the use of powdered activated carbon
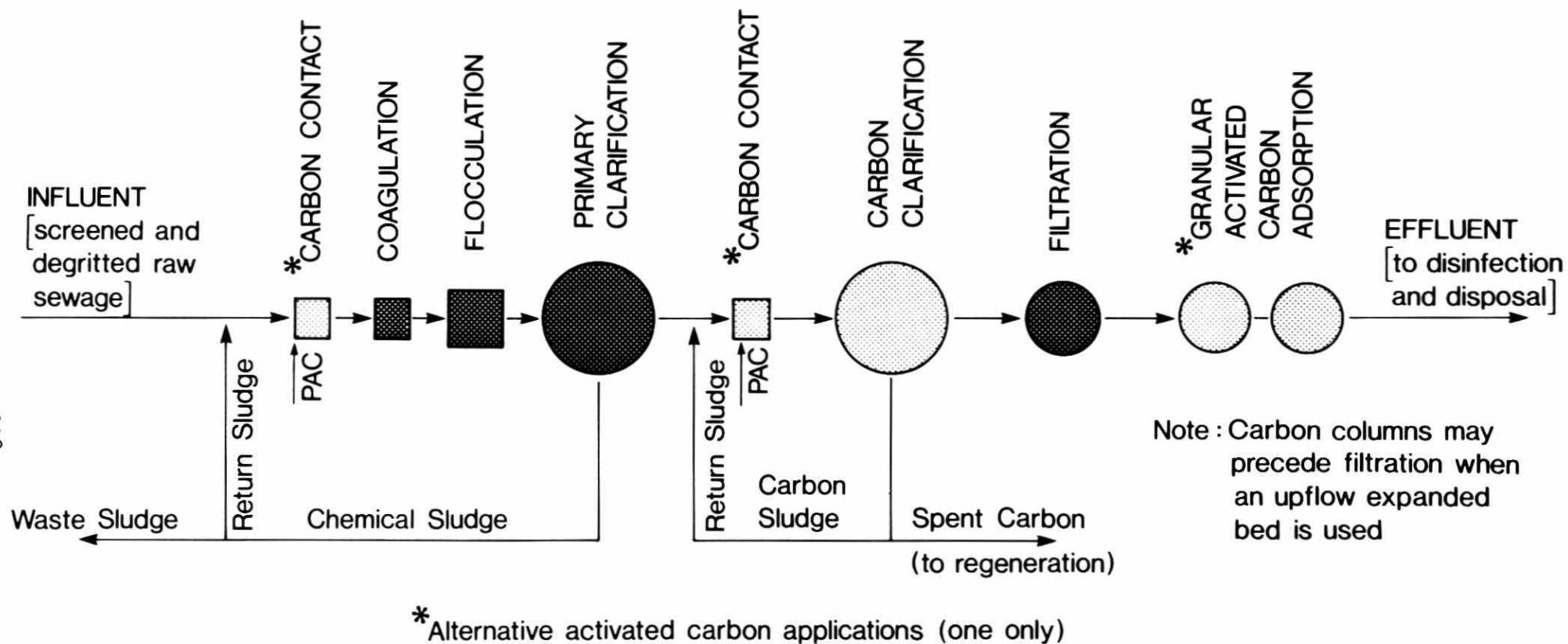
FIGURE 2
ACTIVATED CARBON APPLICATIONS TO PHYSICAL—CHEMICAL PROCESSES

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improved the removal of soluble organic materials (measured as filtered BOD_5 and filtered COD) but decreased the suspended solids removal efficiency. The lower activated carbon dose resulted in slightly better overall performance. Although the mixed liquor containing powdered carbon settled more rapidly and produced a lower sludge volume index (SVI) than the control mixed liquor, the mixed liquor containing the carbon left a greater quantity of residual or non-settleable suspended solids in the plant effluent. In an attempt to reduce the solids carry-over in the carbon treated effluent, an additional experiment was conducted using 25 mg/l powdered activated carbon, 150 mg/l alum and 0.5 mg/l of a cationic polyelectrolyte. The results, shown in Table 3, indicate that the addition of the polymer did not improve the removal of suspended solids and may possibly have hindered the adsorption of soluble organic material.

The results obtained from the three experiments with powdered activated carbon addition to mixed liquor prior to final clarification, at doses of two to three times those estimated for tertiary granular activated carbon application, indicate that powdered activated carbon has very little beneficial effect with respect to overall process performance.

PAC in Physical-Chemical Treatment (Bolton WPCP)

The Ontario Ministry of the Environment conducted full scale experiments with powdered activated carbon in conjunction with chemical clarification at the Bolton Water Pollution Control Plant. There were four experimental runs: 100 mg/l and 200 mg/l PAC (Aqua Nuchar A) with alum and polymer and parallel controls without PAC or chemicals, and 50 mg/l and 200 mg/l PAC with alum and polymer and parallel controls without PAC but with chemicals. The alum dosage was established in laboratory tests to be 200 mg/l. Preliminary full scale tests in-dicated that the use of 1 mg/l of a cationic polyelectrolyte was necessary to prevent activated carbon carry-over in the primary clarifier effluent. As the existing plant geometry did not permit experimentation with sludge recycle, parallel tests with sludge recycle were carried

TABLE 1. PAC ADDITION TO MIXED LIQUOR (BOLTON WPCP)
25 mg/1 PAC & 150 mg/1 ALUM

	BOD_5		COD		SS	VSS	Phosphorus		NO_3
	Total	Filt.	Total	Filt.			Total	Filt.	N
Treated Effluent[1]	6	2	21	14	18	12	0.53	0.07	26
Control Effluent[1]	15	4	34	28	12	7	0.67	0.21	13
Relative Improvement[2]	+60	+50	+38	+50	-50	-71	+21	+67	-100

TABLE 2. PAC ADDITION TO MIXED LIQUOR (BOLTON WPCP)
50 mg/1 PAC & 150 mg/1 ALUM

	BOD_5		COD		SS	VSS	Phosphorus		NO_3
	Total	Filt.	Total	Filt.			Total	Filt.	N
Treated Effluent[1]	15	4	46	18	30	18	0.93	0.19	14
Control Effluent[1]	15	5	29	21	16	9	0.85	0.16	11
Relative Improvement[2]	0	+20	-59	+14	-88	-100	-09	-19	-27

TABLE 3. PAC ADDITION TO MIXED LIQUOR (BOLTON WPCP)
25 mg/1 PAC, 150 mg/1 ALUM & 0.5 mg/1 POLYMER

	BOD_5		COD		SS	VSS	Phosphorus		NO_3
	Total	Filt.	Total	Filt.			Total	Filt.	N
Treated Effluent[1]	16	5	39	17	28	17	0.94	0.10	10
Control Effluent[1]	12	4	27	18	18	9	0.88	0.10	9
Relative Improvement[2]	-33	-25	-44	+06	-56	-89	-07	0	-11

[1] all parameters in mg/1

[2] (control effluent concentration - treated effluent concentration) as a percentage of the control effluent concentration

out in a pilot scale solids-contact type clarifier. Both the conventional primary clarifier [13.9 x 3.0 x 2.4 m (45.5 x 10 x 8 feet)] and the pilot scale solids-contact unit were loaded at 0.55 m^3/m^2h (320 gpd/ft^2) under the intermittant flow conditions created by the method of operation of the raw sewage pumps.

The results of the four experiments are summarized in Tables 4, 5, 6 and 7. Comparison of the control data to that from the PAC additions indicates that the majority of the effluent quality amelioration was attributable to the alum and polymer and not to the activated carbon, the exception being appreciable removal of soluble BOD_5 and TOD at the 50 mg/l PAC dosage. Comparison of the removal efficiencies for the various wastewater parameters with PAC addition (Figure 3) indicates that the removal of suspended parameters was improved at PAC doses up to 100 mg/l but deteriorated thereafter, while the removal of soluble parameters continued to improve with increasing PAC concentration. The limited clarification capacity demonstrated by these results may possibly be improved by the use of larger polyelectrolyte doses. The additional solids contact time provided by sludge recycle in the solids-contact clarifier did not have any significant overall beneficial effects on the process performance. Although the pilot scale unit produced slightly better removals of suspended solids and total phosphorus than the full size conventional clarifier, the removals of the organic parameters were lower. None of the studies with powdered activated carbon addition to raw sewage in combination with coagulants produced an effluent quality suitable for discharge without further treatment.

The use of the powdered activated carbon and coagulants resulted in an increase in sludge concentration relative to the control during the early study periods (when no coagulants were added to the control side of the plant) and the use of PAC caused a slight increase in sludge concentration compared to the coagulant treated control. There was also an increase in sludge pumping (as much as 50%) during the later study periods. Typical average raw sludge concentrations were 8.5% to 9.2% for the carbon treated side of the plant and 6.1% to 8.2% for the control side of the plant.

TABLE 4. PAC ADDITION TO RAW SEWAGE (BOLTON WPCP)
50 mg/1 PAC, 200 mg/1 ALUM & 1.0 mg/1 POLYMER

	BOD$_5$			TOD			SS	Phosphorus	
	Total	Filt.	Susp.	Total	Filt.	Susp.		Total	Filt.
Raw Sewage (mg/1)	90	33	57	208	116	92	173	5.2	2.8
Primary Effluent (control with Alum & Polymer) (mg/1)	59	35	24	158	110	48	78	2.0	0.07
Removal (%)	34	-6	58	24	5	48	55	62	98
Primary Effluent (Carbon treated) (mg/1)	49	25	24	153	89	64	87	1.9	0.13
Removal (%)	46	24	58	26	23	30	50	63	95
Pilot Solids-Contact Clarifier Effluent (mg/1)	57	39	18	166	123	43	55	1.6	0.20
Removal (%)	37	-18	68	20	-6	53	68	69	93

TABLE 5. PAC ADDITION TO RAW SEWAGE (BOLTON WPCP)
100 mg/1 PAC, 200 mg/1 ALUM & 1.0 mg/1 POLYMER

	BOD$_5$			TOD			SS	Phosphorus Total
	Total	Filt.	Susp.	Total	Filt.	Susp.		
Raw Sewage (mg/1)	105	22	83	219	102	117	235	5.2
Primary Effluent (Control without chemicals) (mg/1)	79	37	42	185	121	64	81	4.0
Removal (%)	25	-68	49	16	-19	45	66	23
Primary Effluent (Carbon treated) (mg/1)	34	15	19	107	77	30	64	1.4
Removal (%)	68	32	77	51	25	74	73	73
Pilot Solids-Contact Clarifier Effluent (mg/1)	56	33	23	124	101	23	73	1.2
Removal (%)	47	-50	72	43	1	80	69	77

TABLE 6. PAC ADDITION TO RAW SEWAGE (BOLTON WPCP)
200 mg/l PAC, 200 mg/l ALUM & 1.0 mg/l POLYMER

| | BOD$_5$ | | | TOD | | | SS | Phosphorus |
	Total	Filt.	Susp.	Total	Filt.	Susp.		Total
Raw Sewage (mg/1)	131	47	84	294	193	101	191	5.8
Primary Effluent (control without chemicals) (mg/1)	107	38	69	282	157	125	126	5.6
Removal (%)	18	19	18	4	19	-24	34	3
Primary Effluent (Carbon treated) (mg/1)	41	18	23	115	89	26	59	0.9
Removal (%)	69	62	73	61	54	74	69	84
Pilot Solids-Contact Clarifier Effluent	39	22	17	126	97	29	48	0.6
Removal (%)	70	53	80	57	50	71	75	90

TABLE 7. PAC ADDITION TO RAW SEWAGE (BOLTON WPCP)
200 mg/l PAC, 200 mg/l ALUM & 1.0 mg/l POLYMER

| | BOD$_5$ | | | COD | | | SS | Phosphorus | |
	Total	Filt.	Susp.	Total	Filt.	Susp.		Total	Filt.
Raw Sewage (mg/1)	90	28	62	118	47	71	131	4.7	2.5
Primary Effluent (Control with Alum and Polymer) (mg/1)	52	12	40	56	27	29	65	1.5	0.12
Removal (%)	42	57	35	53	43	59	50	68	95
Primary Effluent (Carbon treated) (mg/1)	69	11	58	72	20	52	79	1.6	0.13
Removal (%)	23	61	6	39	57	27	40	66	95

EFFECT OF PAC ADDITION TO ALUM-COAGULATED PRIMARY CLARIFICATION (Bolton WPCP)

FIGURE 3
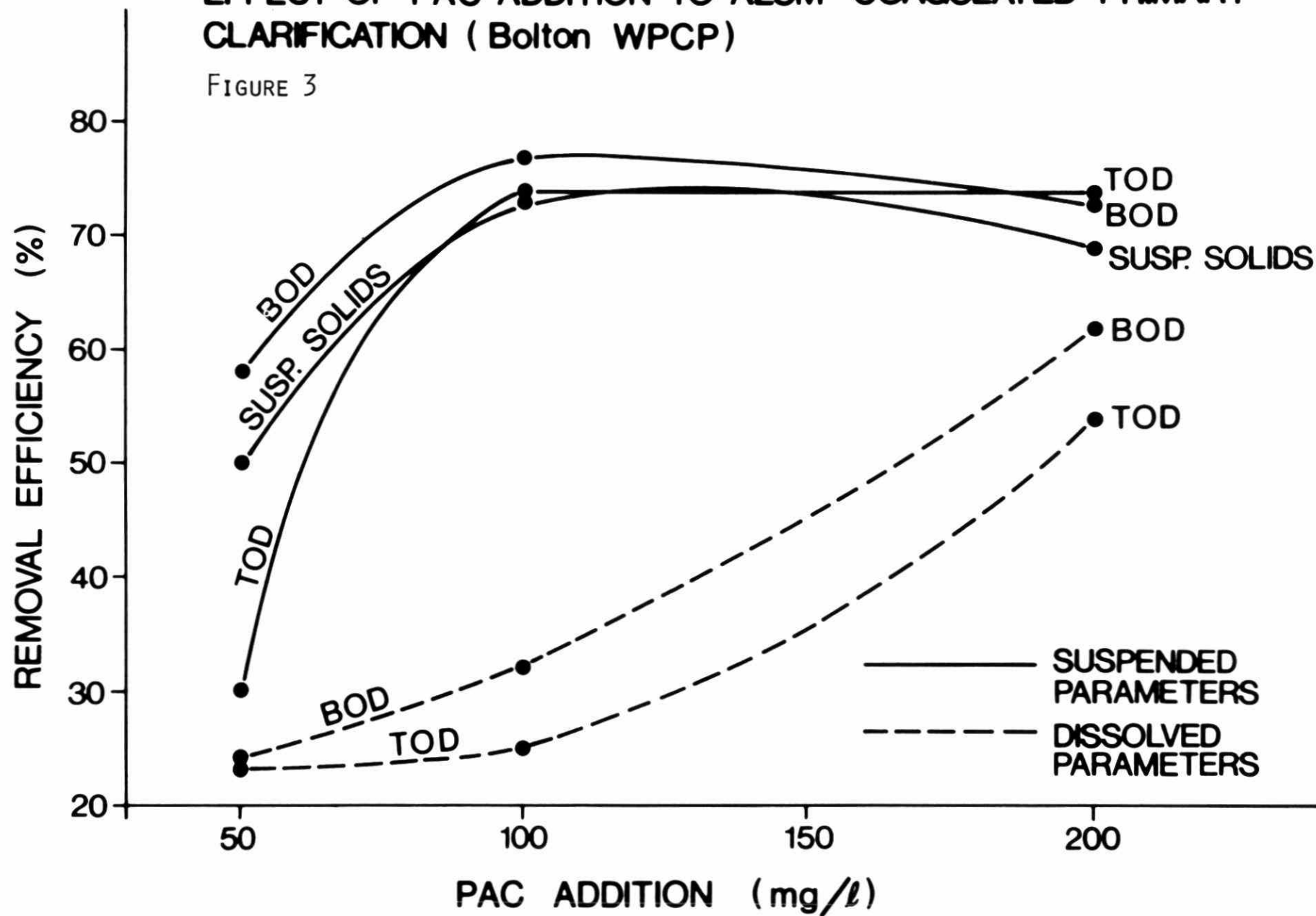
PAC in Physical-Chemical Treatment (Wastewater Technology Centre)

Pilot scale studies of the addition of powdered activated carbon to primary clarification in physical-chemical treatment were undertaken during 1973, 1974 and 1975 at the Wastewater Technology Centre. The results of the first phases of this study were reported earlier (Wei and Le Clair, 1974). Additional phases of the completed study are reported herein. The objectives were to evaluate the performance of powdered activated carbon, when used with chemical coagulation and clarification, in terms of effluent quality, sludge production, clarification efficiency, and carbon utilization efficiency.

The pilot plant at the Wastewater Technology Centre was operated at a constant sewage flow of 5.3 m^3/h (20 gpm). The clarifier had a nominal detention time of two hours and an overflow rate of 1.2 m^3/m^{2}h (600 gpd/ft^2). During experiments with powdered activated carbon addition, a stirred contact tank was placed before the rapid mix tank to provide contact between the activated carbon slurry and raw sewage. For powdered carbon recycle, the sludge was returned to the contact tank at various rates. A flow diagram for the pilot plant is included as Figure 4. The experimental programme included baseline operation without carbon, coagulants or flocculant aids; operation with coagulant alone, and coagulant with a flocculant aid; operation with various activated carbon dosages; and operation with various sludge recycle rates. Standard jar test procedures were used to select the coagulant and flocculant aid dosages which were then held constant throughout the pilot plant experiments. Alum at 150 to 200 mg/l proved to be the optimum coagulant and an anionic polymer at 0.50 mg/l was selected as the flocculant aid.

Seven experimental runs were undertaken between June 1973 and May 1974 to investigate the effects of alum, polymer, powdered activated carbon (Aqua Nuchar) and sludge recycle on effluent quality and sludge characteristics. The results of these experiments are reported elsewhere (Wei and Le Clair, 1974) but may be summarized as follows. The use of alum as a coagulant significantly improved clarification as compared to baseline efficiency (no chemicals). The removal of total BOD$_5$, soluble TOC, suspended solids, total phosphorus,

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DETENTION TIMES AT 20 Igpm

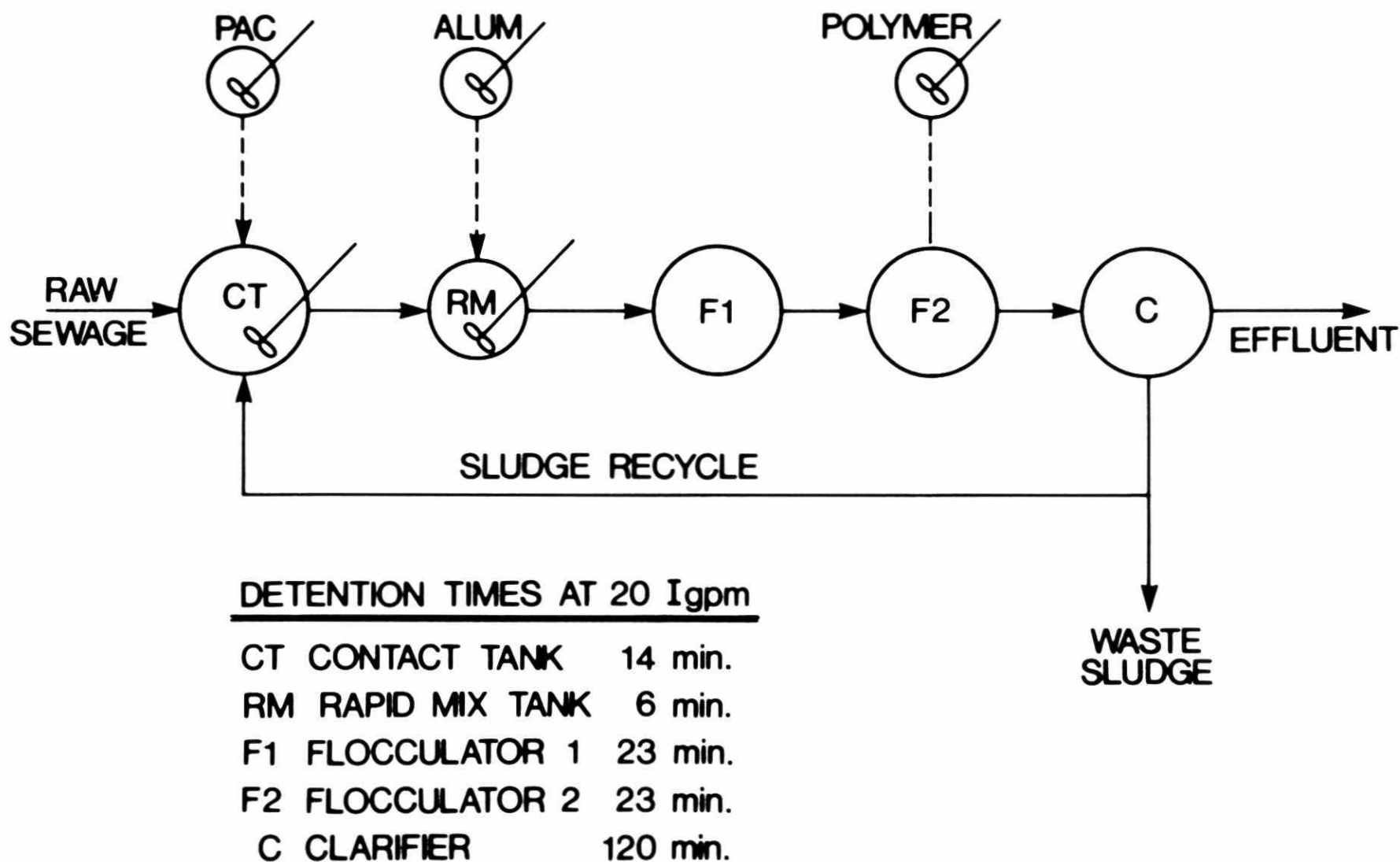|		
CT	CONTACT TANK	14 min.
RM	RAPID MIX TANK	6 min.
F1	FLOCCULATOR 1	23 min.
F2	FLOCCULATOR 2	23 min.
C	CLARIFIER	120 min.

FIGURE 4
PAC PILOT PLANT FLOW DIAGRAM (Wastewater Technology Centre)

copper, zinc, and iron were significantly increased while the concentration of aluminum in the plant effluent increased as a result of the carry-over of aluminum hydroxide flocs with the clarifier effluent suspended solids. The addition of polymer caused a small but significant decrease in the effluent BOD_5, TOC, and suspended solids concentrations. The addition of powdered activated carbon at 200 mg/l without sludge recycle resulted in small and statistically insignificant improvements in total BOD_5 and soluble TOC removals. However, the use of PAC did improve clarification, presumably by increasing the weight of flocs; this improvement was indicated by significant changes in effluent suspended solids and phosphorus concentration and was confirmed by column settling tests which showed faster settling rates with PAC addition.

It was expected that the adsorptive capacity of the powdered activated carbon was not being utilized effectively under one-pass operational conditions and that both the carbon utilization and effluent quality may be improved by recycling the sludge to the plant influent. To test this hypothesis, further pilot scale experiments were performed. This portion of the PAC experimental programme was carried out between September 1974 and August 1975. Nine experimental runs included combinations of five PAC doses (0, 25, 50, 100 and 200 mg/l), and three sludge recycle rates (0, 5 and 15% by volume). The alum and polymer doses were held constant at 150 and 0.50 mg/l respectively for all experiments in this series.

The influent and effluent characteristics are presented in Tables 8 and 9 with the respective removal efficiencies listed in Table 10. Response surface plots of total COD, total BOD_5, suspended solids and soluble TOC removals are included as Figures 5, 6, 7 and 8. These plots indicate that the removal of total parameters was optimized at an activated carbon dose of 100 mg/l and a sludge recycle rate of 5%. The removal of filtered TOC, the only soluble parameter monitored, indicates that the optimum recycle rate for removal of soluble organics was also 5%, but that removal efficiency increased with increasing PAC concentration up to 200 mg/l, the largest dose studied.

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TABLE 8. PAC ADDITION TO RAW SEWAGE (WTC)
RAW SEWAGE CHARACTERISTICS

PAC mg/l	SLUDGE RECYCLE %	TOTAL BOD$_5$	TOTAL COD	FILT. TOC	TOTAL P	TOTAL Al	SUSPENDED SOLIDS
0	0	128	350	32	9.1	4.2	286
100	0	133	331	29	6.5	0.2	230
0	5	108	416	33	8.0	3.1	241
50	5	153	473	37	13.5	3.6	421
100	5	162	633	35	13.0	4.0	448
200	5	152	409	30	11.3	3.4	345
25	15	121	348	30	7.8	2.2	224
50	15	120	368	31	8.1	2.6	343
200	15	138	466	26	9.0	3.7	345

. All data in mg/l

. Alum: 150 mg/l

. Polymer: 0.5 mg/l

TABLE 9. PAC ADDITION TO RAW SEWAGE (WTC)
         EFFLUENT CHARACTERISTICS

PAC mg/l	SLUDGE RECYCLE %	TOTAL BOD$_5$	TOTAL COD	FILT. TOC	TOTAL P	TOTAL Al	SUSPENDED SOLIDS
0	0	21	84	19	2.2	1.0	34
100	0	17	76	20	0.8	0.7	20
0	5	17	67	19	0.9	0.5	15
50	5	14	54	20	1.0	0.6	21
100	5	13	54	17	1.3	0.7	12
200	5	16	87	17	0.5	0.7	19
25	15	19	67	27	1.6	0.6	12
50	15	11	62	16	0.5	0.6	13
200	15	16	75	15	0.5	0.8	25

. All data in mg/l

. Alum: 150 mg/l

. Polymer: 0.5 mg/l

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TABLE 10. PAC ADDITION TO RAW SEWAGE (WTC)
         PERCENT REMOVALS

PAC mg/l	SLUDGE RECYCLE %	TOTAL BOD$_5$	TOTAL COD	FILT. TOC	TOTAL P	TOTAL Al	SUSPENDED SOLIDS
0	0	83	74	37	76	73	88
100	0	86	77	27	88	-276	91
0	5	88	83	33	91	85	95
50	5	90	88	44	91	84	95
100	5	89	89	44	90	84	98
200	5	85	81	47	93	64	91
25	15	85	81	8	75	31	94
50	15	90	82	38	94	75	96
200	15	88	84	42	94	77	92

. All data in mg/l

. Alum: 150 mg/l

. Polymer: 0.5 mg/l

SLUDGE RECYCLE RATE (percent of wastewater flow)

PAC ADDITION (mg/ℓ)
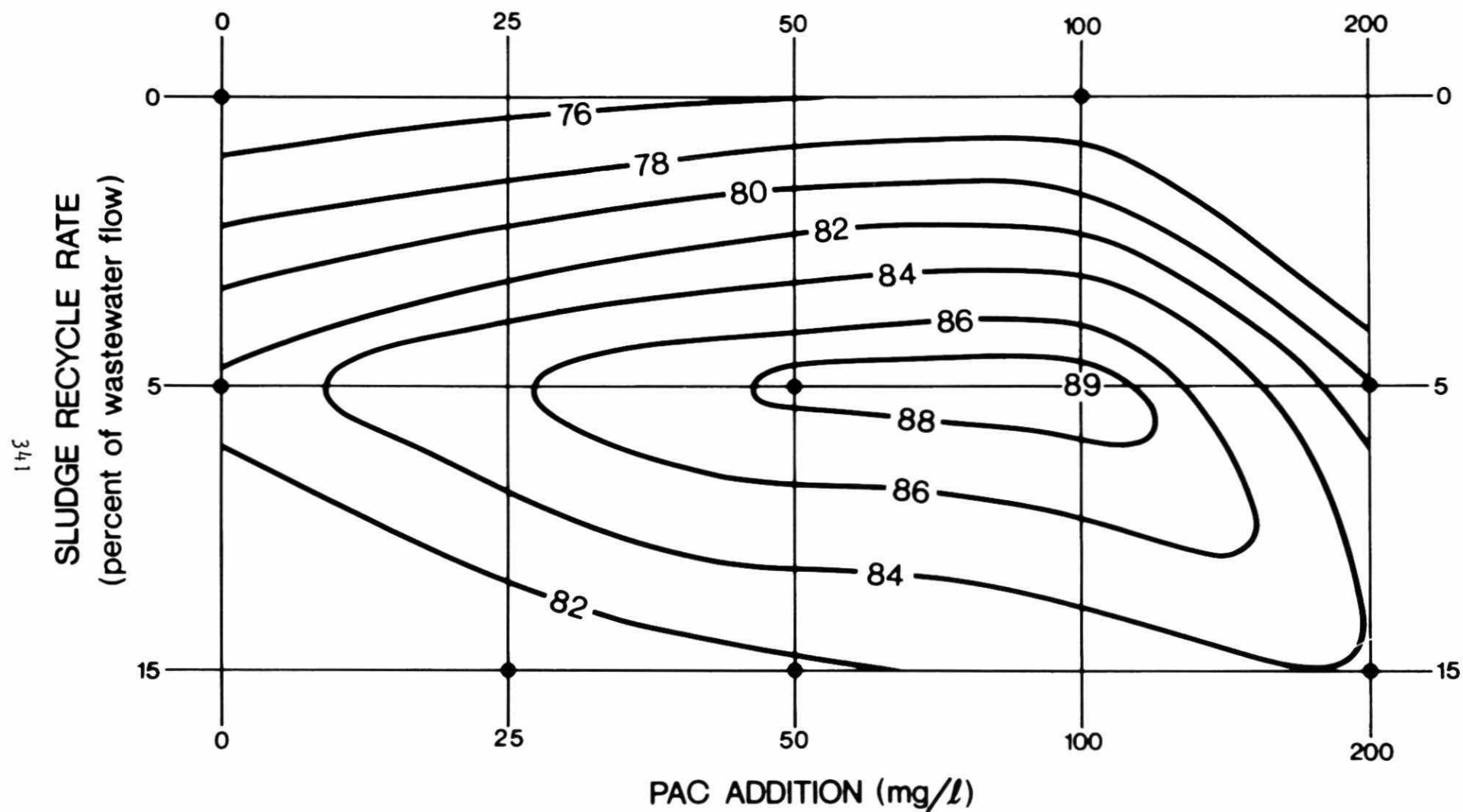
● INDICATES DATA POINT EXCEPT WHERE REMOVAL VALUE IS SHOWN

FIGURE 5   TOTAL COD PERCENT REMOVALS

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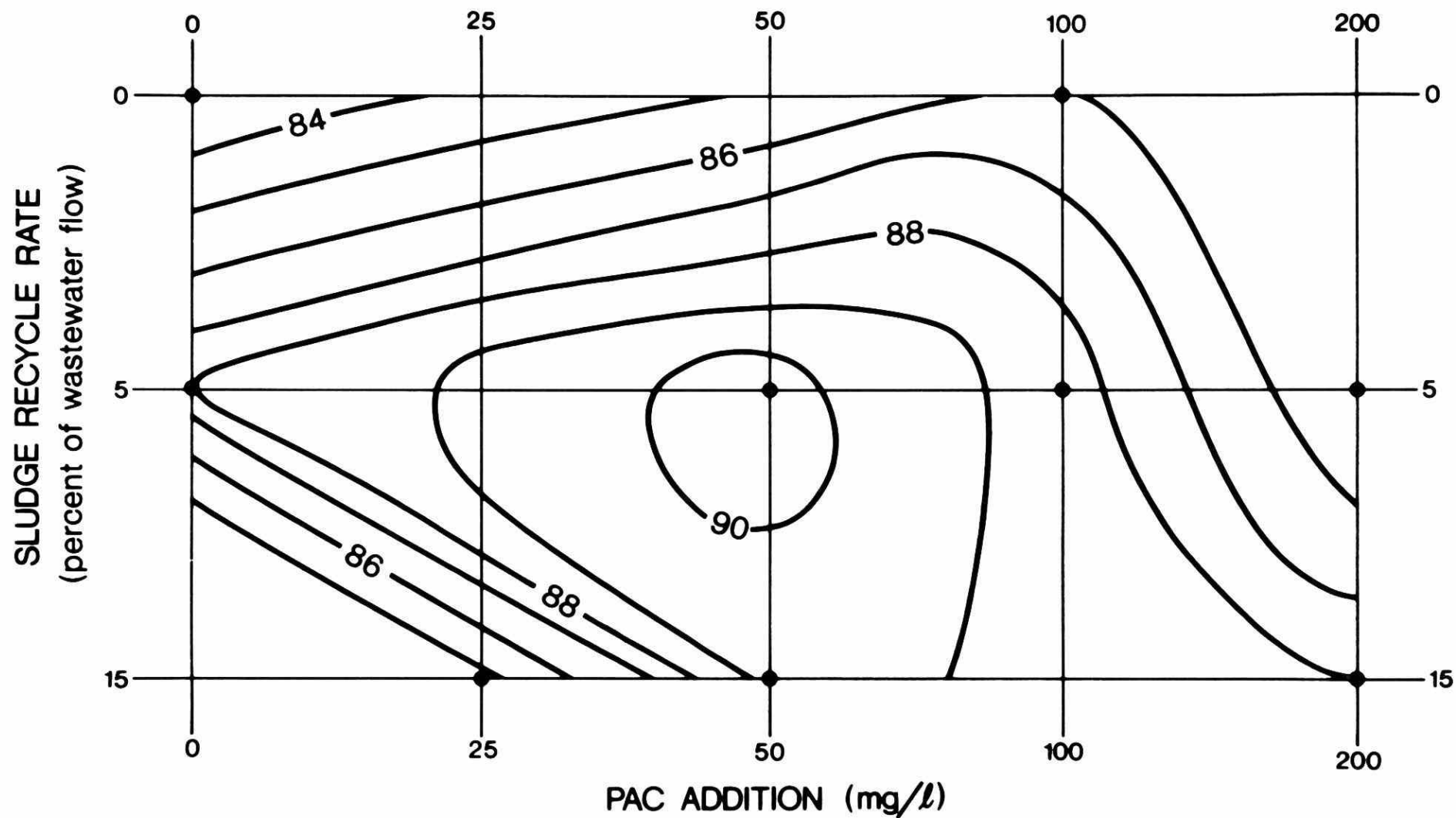
● INDICATES DATA POINT EXCEPT WHERE REMOVAL VALUE IS SHOWN

FIGURE 6   TOTAL BOD₅ PERCENT REMOVALS

● INDICATES DATA POINT EXCEPT WHERE REMOVAL VALUE IS SHOWN

FIGURE 7 SUSPENDED SOLIDS PERCENT REMOVALS

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FIGURE 8 SOLUBLE TOC PERCENT REMOVALS

● INDICATES DATA POINT EXCEPT WHERE REMOVAL VALUE IS SHOWN

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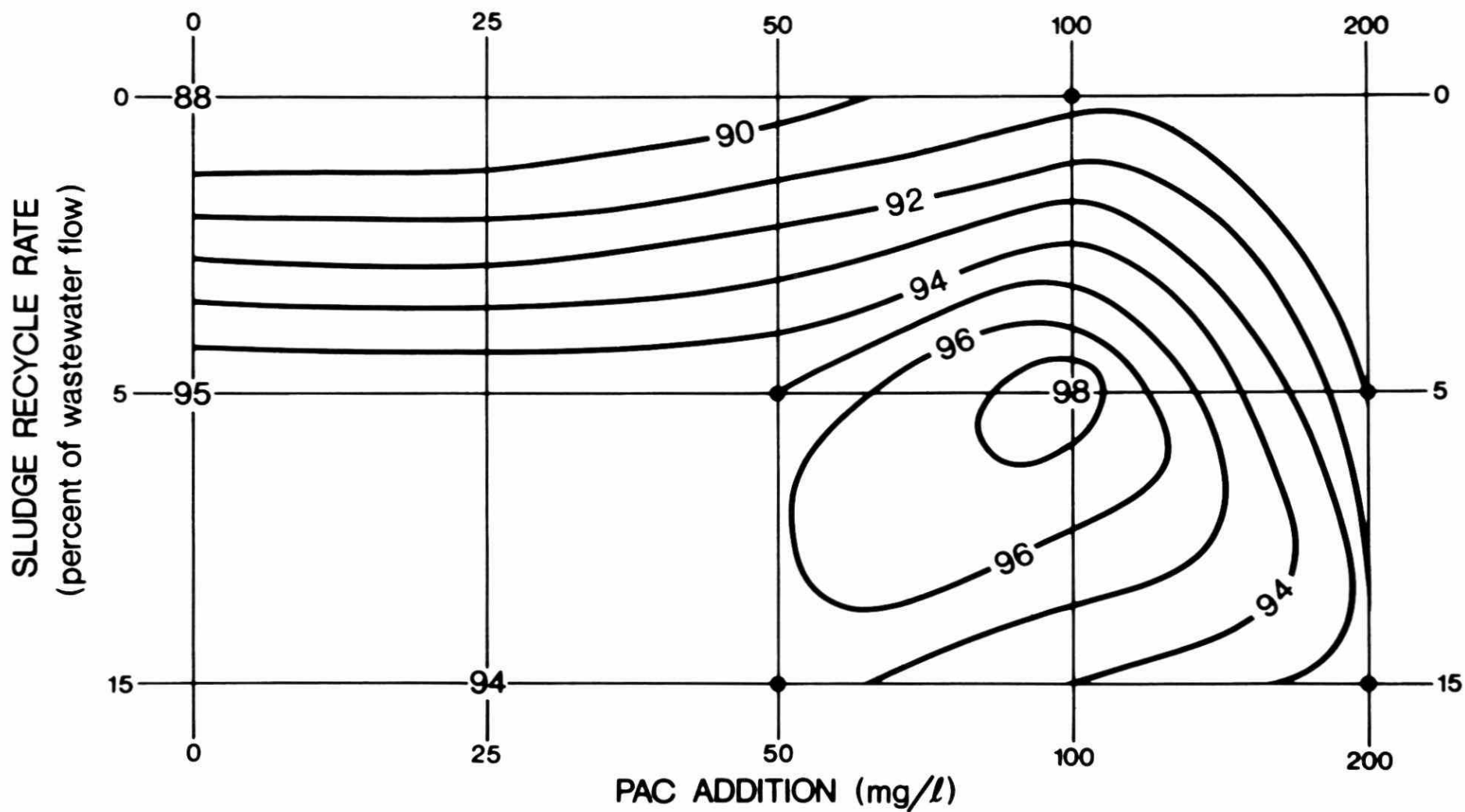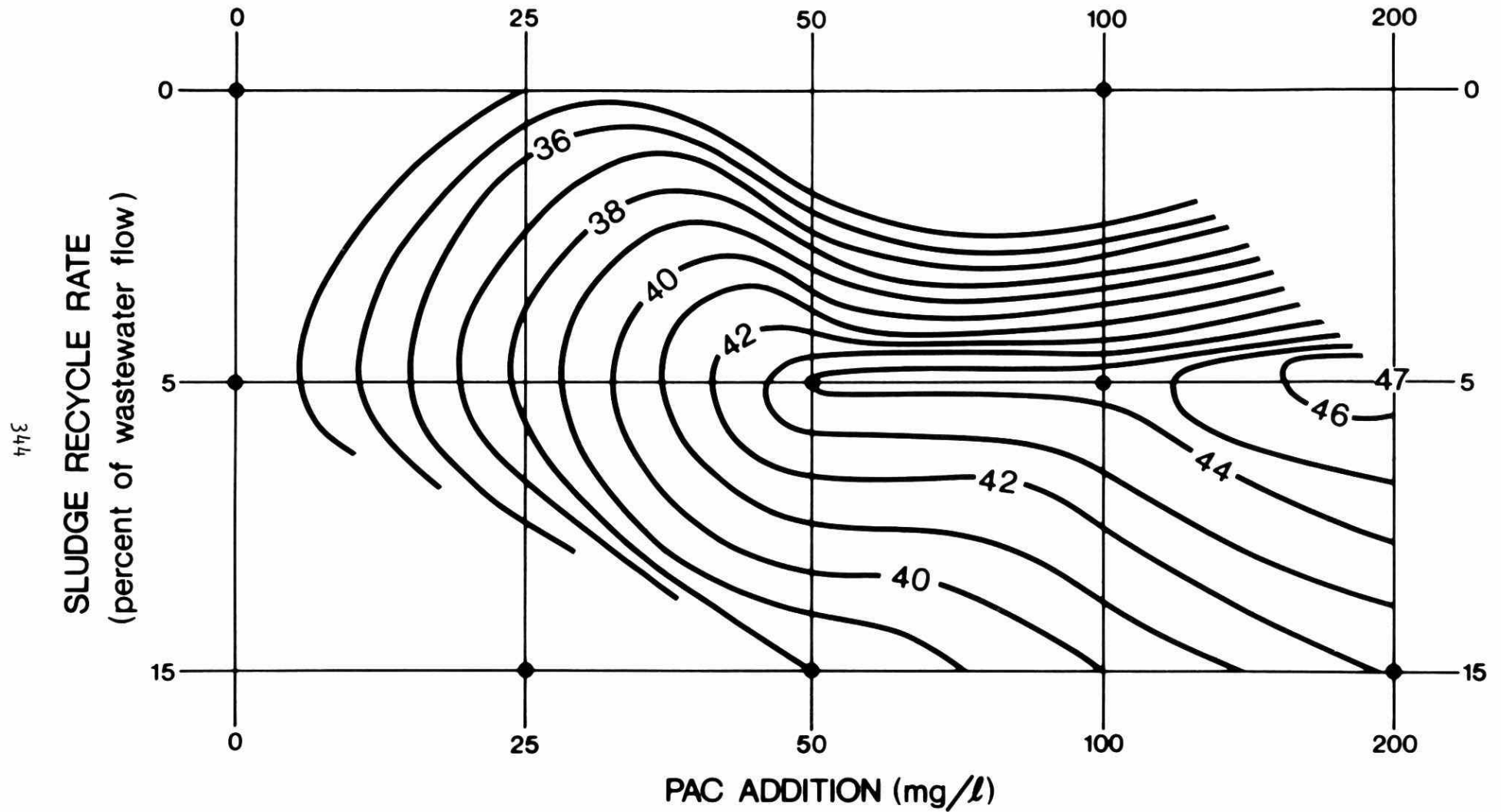Under the optimum operating conditions of 100 mg/l PAC addition and 5% sludge recycle rate, with 150 mg/l alum as a coagulant and 0.5 mg/l of anionic polymer as a flocculation aid, the following average effluent concentrations and removals were obtained: 13 mg/l (89%) total BOD_5, 54 mg/l (89%) total COD, 17 mg/l (44%) soluble TOC, 1.3 mg/l (90%) total phosphorus, 12 mg/l (98%) suspended solids. The average sludge concentration produced by the above operating conditions was 2.37% suspended solids.

Certain trends were observed in activated carbon adsorptive capacity utilization under the operational conditions applied. If one considers the average concentration of soluble TOC removed per unit of PAC applied (mg/l TOC per mg/l PAC), the effects of both PAC addition and sludge recycle rate may be identified. At the 100 mg/l PAC dosage the quantity of TOC adsorbed increased from 0.09 gm/gm at 0% recycle to 0.19 gm/gm at 5% recycle. This increase in effective adsorption capacity was caused in part by an increase in the soluble TOC content of the raw sewage between the 0% and 5% recycle experiments; however, the magnitude of the increase suggests that the use of sludge recycle was a significant factor. Increasing the recycle rate to 15% slightly decreased the amount of TOC adsorption per unit weight of carbon, although this effect probably resulted in part from the lower TOC concentrations in the raw sewage during the 15% recycle runs. The average amount of TOC adsorbed per unit weight of carbon decreased as the carbon dose increased as expected. This same effect may be the reason for the limited improvements obtained by increasing the sludge recycle rate from 5 to 15%. In summary, both the effluent quality and carbon utilization were optimized at a sludge recycle rate of 5%; the effluent quality was optimized at a PAC dose of 100 mg/l, but the lowest carbon dose favoured maximum utilization of the carbon's adsorptive capacity.

Discussion - PAC

The experiments on the addition of powdered activated carbon to primary clarification undertaken by the Ontario Ministry of the

Environment at the Bolton WPCP and by Environment Canada at the Wastewater Technology Centre have demonstrated the potential of this mode of activated carbon application and the effects of varying the PAC dose and sludge recycle rates. In both sets of experiments it was demonstrated that the removal of soluble organic substances is increased with increasing PAC addition up to and possibly beyond 200 mg/l. Also, the removal of suspended solids and the total wastewater parameters (BOD_5, COD, TOD) is maximized at a PAC dose of approximately 100 mg/l; PAC addition at rates in excess of 100 mg/l resulted in slight decreases in the removal efficiency for these parameters.

The Wastewater Technology Centre experiments have demonstrated that recycling clarifier sludge to the PAC contact tank produces improved removal of all wastewater parameters up to a sludge recycle rate of approximately 5%, but causes slightly reduced efficiencies when the rate is increased to 15% of the wastewater flow. The value of recycling sludge was not confirmed by the Bolton experiment with a solids-contact type clarifier; however, any comparison between a solids-contact type unit and a conventional rectangular clarifier would be significantly influenced by the differences in the hydraulic conditions inherent within the units even when the same surface loading rate is applied to both units.

It may be concluded from these studies that the use of powdered activated carbon in primary clarification is not a practical process for general application. In a study using an existing full size plant a low quality effluent was produced; the relatively low efficiency was caused in part by the lack of a sludge recycle system, although the extent to which sludge recycle may have improved the process performance is currently unknown. The operating costs were high as a result of activated carbon consumption since there is no practical process available for the separation of spent PAC from primary sludge to facilitate carbon regeneration and re-use.

The PAC experiments described above involve the addition of carbon to raw wastewater prior to coagulant addition and with the use of one clarification step to remove suspended solids and spent activated

carbon. An alternative approach is to apply the PAC after primary chemical coagulation and clarification; this method has been investigated elsewhere. A study by Burns and Shell (1973) described the addition of alum, ferric chloride and lime to raw domestic sewage followed by PAC addition and granular media filtration. The pilot plant consisted of three solids-contact clarifiers and one granular media filter. Of the three clarifiers, one was used for chemical coagulation and the remaining two for addition of PAC in a counter-current sequence. Experiments were also carried out using one clarifier for single-stage powdered carbon contacting. In another pilot scale experiment Beebe (1973) studied the addition of alum and polymer to raw sewage in a solids-contact clarifier followed by the addition of PAC in a second solids-contact clarifier. The results of these experiments indicate that there is little or no benefit to be realized in terms of effluent quality by separating PAC addition from chemical clarification. However, carbon re-use would be greatly simplified by maintaining separate carbon and sludge streams, providing, of course, that an economical method of carbon regeneration is available.

The study of powdered activated carbon addition prior to secondary clarification in the activated sludge process resulted in little overall improvement in effluent quality. The study included carbon doses up to 50 mg/l in conjunction with coagulants. The spent PAC was mixed with the activated sludge; therefore, this process would also suffer from high operational costs as a result of the apparent lack of a feasible PAC separation and regeneration process.

GRANULAR ACTIVATED CARBON

GAC in Tertiary Treatment (Bolton WPCP)

At the Bolton Water Pollution Control Plant, studies were also carried out on granular activated carbon for the treatment of secondary effluents. A pilot scale carbon contact process was added to the full scale conventional activated sludge treatment plant for this study. The pilot scale apparatus consisted of four 18 inch

diameter columns operated in series in the downflow mode. The entire system was backwashed manually once every 24 hours of operating time. The performance of the carbon columns was monitored principally on the basis of filtered COD. The initial soluble COD concentration in the carbon column effluent was approximately 6 mg/l; the effluent quality deteriorated slowly to soluble COD's of approximately 11 mg/l after approximately 100 days. At that time the performance of the system appeared to have reached an equilibrium condition where the effluent quality from each column remained relatively constant.

After 6040 m^3 (1,600,000 gallons) had been treated (130 days) the carbon in the lead column was replaced. However, at this time no distinct breakthrough of COD had been observed through that column. Operation of the adsorption units was then continued for a total of 237 operating days or 11,000 bed volumes [10960 m^3 (2.9 million gallons)] of treated secondary effluent. After the replacement of the carbon in the lead column the effluent soluble COD concentration was approximately 6 to 8 mg/l, but after the first column had treated 6040 m^3 (and columns 2 to 4, 10960 m^3) the effluent quality deteriorated to 8 to 10 mg/l soluble COD. Table 11 presents a summary of the operational efficiency of the pilot scale adsorption units before and after the carbon in the lead column was replaced. Since none of the carbon had been utilized to its ultimate capacity and since the adsorption process had not reached its effective operational capacity (the effluent was still acceptable after 10960 m^3), the ultimate and effective capacities of the carbon could not be determined. It was observed, however, that at volumes approaching 6000 m^3 the process was less capable of adsorbing shock organic loads than when the carbon was relatively fresh. As more of the adsorptive capacity of the carbon is utilized the adsorptive process becomes more sensitive to organic loading. This characteristic will result in unpredictable performance where shock loads are frequent and it may be necessary to change the carbon before its effective capacity has been utilized.

TABLE 11. GAC IN TERTIARY TREATMENT (BOLTON WPCP)

	pH	BOD₅ Total	Filt.	COD Total	Filt.	SS	VSS	Phosphorus Total	Filt.	NO₃ N
Before Lead Column Replaced										
(0 to 1.6 M.G.)										
Secondary Eff.	7.4	15	4	40	27	22	16	1.5	0.92	13
Col. #1 Eff.	7.5	9	3	23	20	6	5	1.1	0.85	13
Col. #2 Eff.	7.5	-	-	-	18	-	-	-	-	-
Col. #3 Eff.	-	-	-	-	14	-	-	-	-	-
Col. #4 Eff.	-	9	-	13	11	3	2	1.1	0.84	13
Overall % Rem.	-	·40	-	68	59	86	88	27	9	0
After Lead Column Replaced										
(1.6 to 2.9 M.G.)										
Secondary Eff.	7.2	15	4	32	22	18	11	1.0	0.22	14
Col. #1 Eff.	7.2	7	3	20	19	4	3	0.5	0.19	14
Col. #2 Eff.	7.2	-	-	-	16	-	-	-	-	-
Col. #3 Eff.	7.4	-	-	-	13	-	-	-	-	-
Col. #4 Eff.	7.4	5	-	9	9	3	-	0.3	0.18	14
Overall % Rem.	-	67	-	72	59	83	-	66	18	0

· All data in mg/l except % Removal

GAC in Physical-Chemical Treatment (Newmarket WPCP)

The Ontario Ministry of the Environment completed a study of granular activated carbon adsorption applied to a chemically clarified primary effluent at the Newmarket Water Pollution Control Plant (Hraseova, 1973). The Newmarket WPCP is a 2.0 MGD conventional activated sludge plant incorporating phosphorus removal in the primary clarifier with 200 mg/l hydrated lime to approximately pH 9.5. Three five inch diameter columns, containing a total of 3.38 m (11.1 ft.) [18.2 kg (40 pounds)] of 12 x 40 granular activated carbon, were applied in series and downflow mode to sand filtered primary clarifier effluent. The carbon columns were loaded at a rate of 0.003 m^3/m^2h (4 gpm/ft^2) and backwashed when the total headloss reached 34.5 kN/m^2 (5 psi).

The activated carbon treated approximately 5,300 bed volumes [189 m^3 (50,000 gallons)] of clarified and filtered wastewater over a period of 86 days. After 900 hours of operation, or 103 m^3 (27,360 gallons) of throughput, the effluent TOC concentration was observed to be 61% of the influent value (Figure 9). Although there was considerable variation in removal efficiency, Figure 9 demonstrates a general tendency for the effluent quality to deteriorate gradually as the total quantity of treated wastewater increased. No distinct breakthrough was observed; the 61% TOC remaining level may be arbitrarily defined as the breakpoint for illustrative purposes, and on that basis the effective adsorptive capacities of the activated carbon for the various parameters were determined (Table 12). Plots of Percent Remaining versus Volume Treated for total COD, BOD_5 and TOD were similar to that for TOC.

The average influent and effluent concentrations and removal efficiencies are presented in Table 13. The high pH of the primary effluent being treated was assumed to be responsible for the limited adsorption capabilities of the process. Suspended solids and total phosphorus removals varied widely, but did not demonstrate a tendency toward reduced removals with increasing time or throughput. Phosphorus removal was attributed to the filtration of the calcium phosphate precipitate by the carbon bed.

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FIGURE 9
TOC BREAKTHROUGH
DOWNFLOW GRANULAR
ACTIVATED CARBON
(Newmarket WPCP)

COAGULANT : LIME
PRE-FILTRATION : SAND
pH : APPROX. 9.5

TOC (% remaining)

VOLUME TREATED (Imp. Gallons x 1000)

351
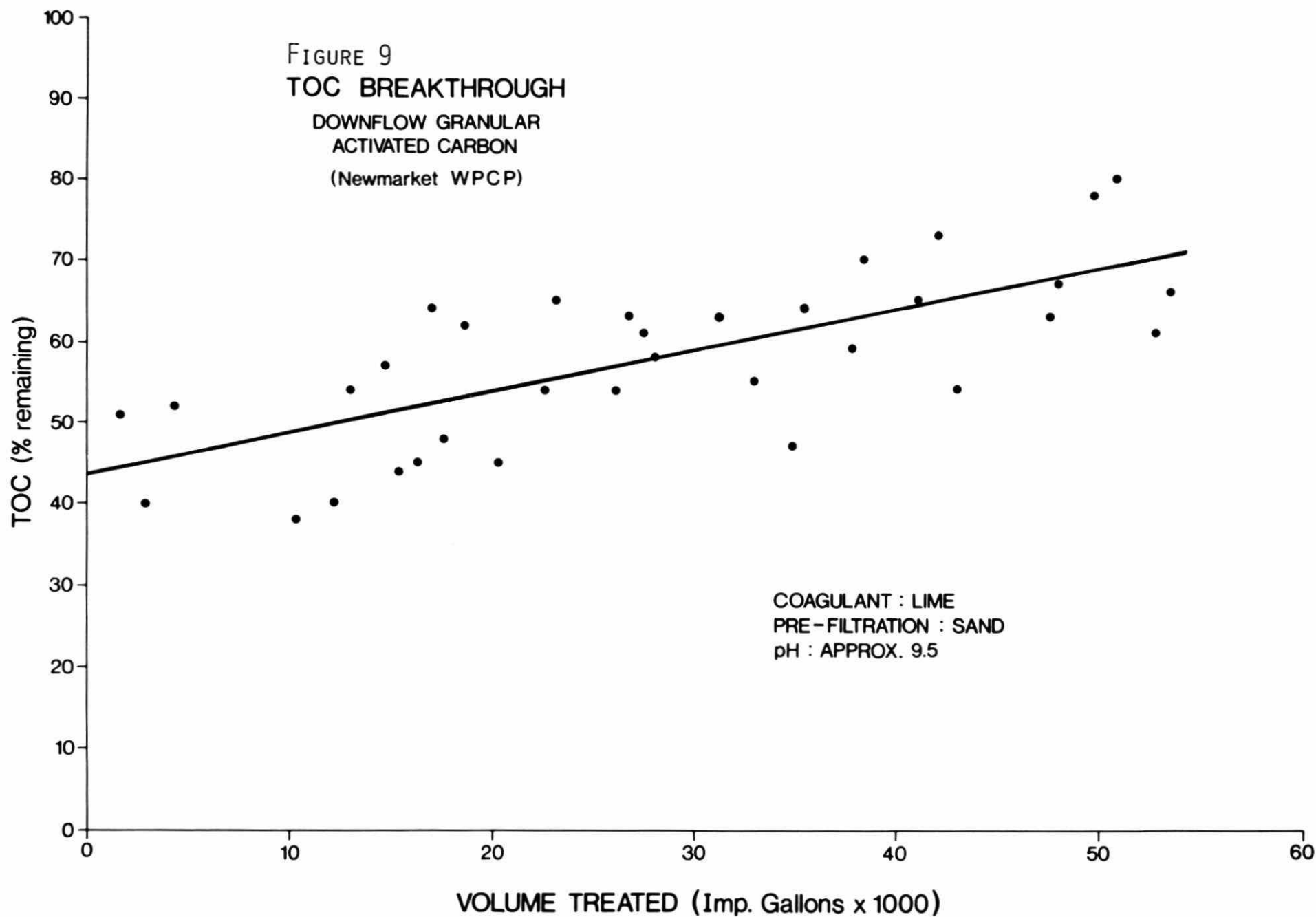
TABLE 12. GAC IN PHYSICAL-CHEMICAL TREATMENT (NEWMARKET WPCP)
ADSORPTION AFTER 900 HOURS (27,360 IMP. GALLONS)

	gm./gm. GAC	PERCENT REMAINING
Total TOC	0.125	61
Total COD	0.7	40
Total BOD$_5$	0.307	44
Total TOD	0.628	60

TABLE 13. GAC IN PHYSICAL-CHEMICAL TREATMENT (NEWMARKET WPCP)
AVERAGE PROCESS PERFORMANCE

	TOTAL BOD$_5$	TOTAL COD	TOTAL TOC	SS	TOTAL P
Column Influent	80	172	46	106	4.0
Column Effluent	35	69	26	23	2.2
Percent Removal	56	60	43	78	45

. All parameters in mg/l except percent removal

. Clarification with 200 mg/l lime, pH approx. 9.5

. Carbon column influent sand filtered

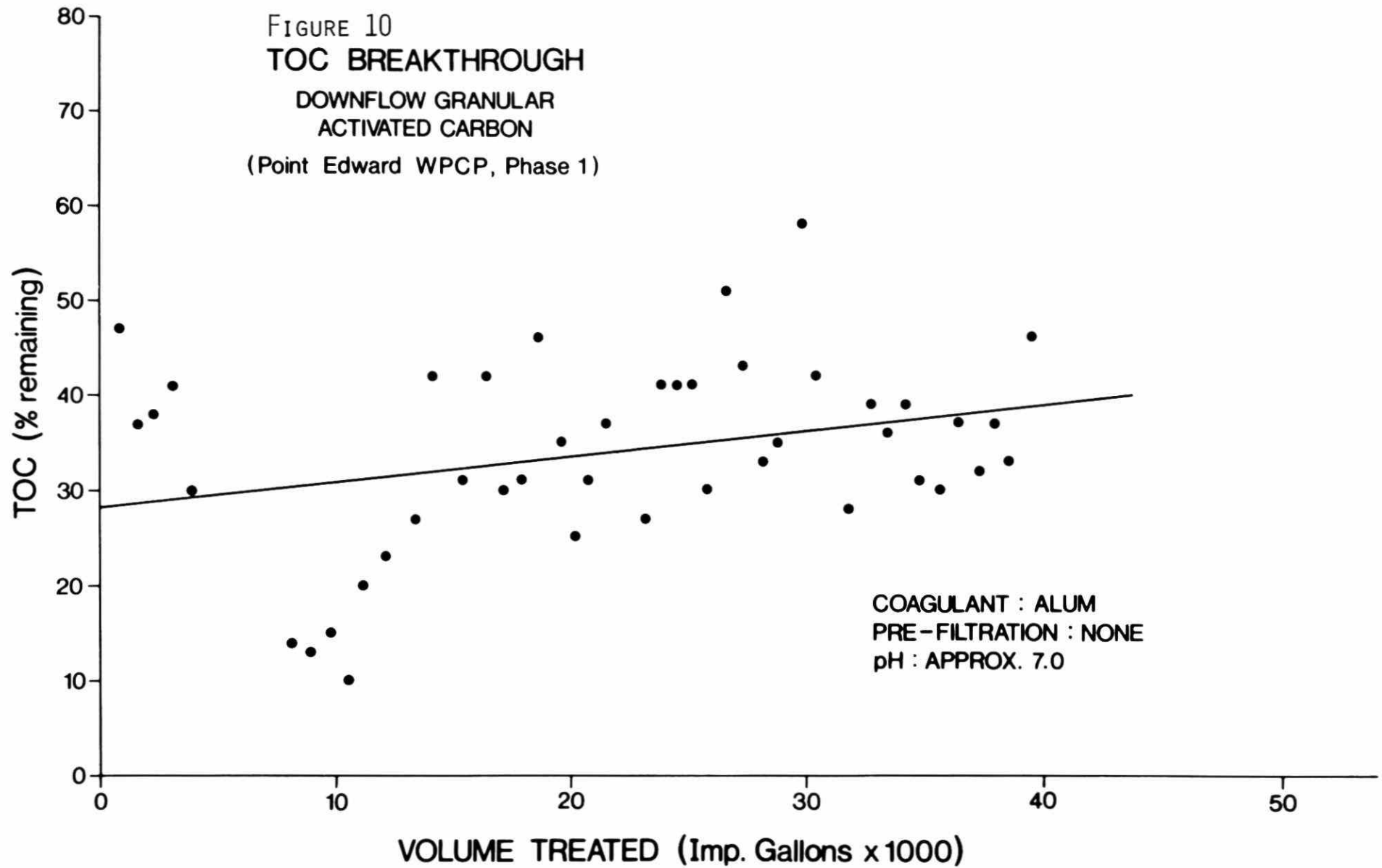## GAC in Physical-Chemical Treatment (Point Edward WPCP)

At the Point Edward Water Pollution Control Plant, the Ontario Ministry of the Environment applied granular activated carbon to both lime and alum coagulated and clarified wastewaters. The Point Edward plant consists of a primary treatment process with discharge to the St. Clair River.

Phase 1 of the carbon adsorption experiments consisted of applying primary clarifier effluent to four five-inch diameter downflow carbon columns connected in series. The carbon size was 12 x 40 mesh, the total carbon depth 3.9 m (12.75 feet), the total carbon weight 20.7 kg (45.59 pounds), and the loading rate 0.003 m^3/m^2h (4 gpm/ft^2). The feed was coagulated with 300 mg/l of alum and flocculated with 0.25 mg/l of an anionic polymer. Backwash was performed manually when a headloss of 34.5 kN/m^2 (5 psi) was reached. The columns were back-washed in parallel for 30 minutes at an upflow rate of 0.005 m^3/m^2h (7 gpm/ft^2).

The average values for wastewater parameters and percent removals over this phase of the work are presented in Table 14. Removal efficiencies demonstrated considerable variability but a slight trend toward reduced removal capabilities can be identified as shown in Figure 10 for total TOC residuals. By the end of the experiment (3,600 bed volumes, 180 m^3, 40,000 gallons) the first column appeared to be nearing exhaustion since its TOD removal capacity had decreased to 10%. At this point the total TOD removal was equivalent to 0.57 gm/gm on column 1, 0.25 gm/gm on column 2, 0.21 on column 3 and 0.15 on column 4. Since total TOD was being measured, some of this removal capacity must be attributed to filtration of organic solids by the carbon bed. Also, biological activity within the bed may have con-tributed to the TOD removal. Influent suspended solids during this phase varied in concentration from 21 to 116 mg/l with an average of 67 mg/l. The pilot scale study demonstrated that the carbon adsorption process could cope with such a solids load when backwashed with 4.5% of the process throughput volume at a rate of 0.005 m^3/m^2h (7 gpm/ft^2).

Phase 2 of the experiments consisted of applying filtered primary effluent to four downflow carbon columns and four upflow

FIGURE 10
TOC BREAKTHROUGH
DOWNFLOW GRANULAR
ACTIVATED CARBON

(Point Edward WPCP, Phase 1)

COAGULANT : ALUM
PRE-FILTRATION : NONE
pH : APPROX. 7.0

TOC (% remaining)

VOLUME TREATED (Imp. Gallons x 1000)

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carbon columns. All columns were five inches in diameter, seven feet high, and connected in series. For the downflow (fixed bed) experiment 4.3 m (14.0 feet) [22.7 kg (50 pounds)] of 12 x 40 mesh size carbon were used. The upflow (expanded bed) columns were charged with a total of 19 kg (41.7 pounds) of 8 x 30 mesh size carbon. Feed to the carbon adsorption units was coagulated with 120 mg/l lime and 20 mg/l ferric chloride, flocculated, clarified, and filtered in a dual media unit. The loading rate for both sets of columns was 0.003 m^3/m^2h (4 gpm/ft^2).

Influent and effluent concentrations and removal efficiencies for both fixed and expanded bed experiments are presented in Table 15. As in Phase 1, the carbon was not utilized to exhaustion and the removal efficiencies demonstrated a gradual decline throughout the run (Figure 11). Approximately 208 m^3 (55,000 gallons) of wastewater had been treated by each set of carbon columns at the completion of the experiment (4,600 bed volumes downflow, 5,500 packed bed volumes upflow).

Comparison of the expanded and fixed bed performances indicates that the fixed bed produced slightly higher percent removals and better effluent quality than the expanded bed. The filtration provided by the fixed bed was probably responsible for the additional removal capacity. The use of air scour prior to backwash, and pre-filtration, permitted a reduction of the backwash water quantity for the fixed bed columns to 0.6% of the treated water volume. However, the headloss over the fixed beds increased steadily and frequent backwashing was required; flushing of the expanded beds was required less frequently.

Comparison of the Phase 2 fixed bed performance to that observed in Phase 1 indicates that the Phase 1 columns produced a higher quality effluent and greater percent removals. The higher percent removals may be attributed in part to the higher concentration of suspended solids in the unfiltered influent in Phase 1. However, both phases produced approximately the same effluent suspended solids concentrations. The difference in effluent quality is therefore mostly attributable to the difference in adsorption capacities for soluble materials between the two experiments. The difference in
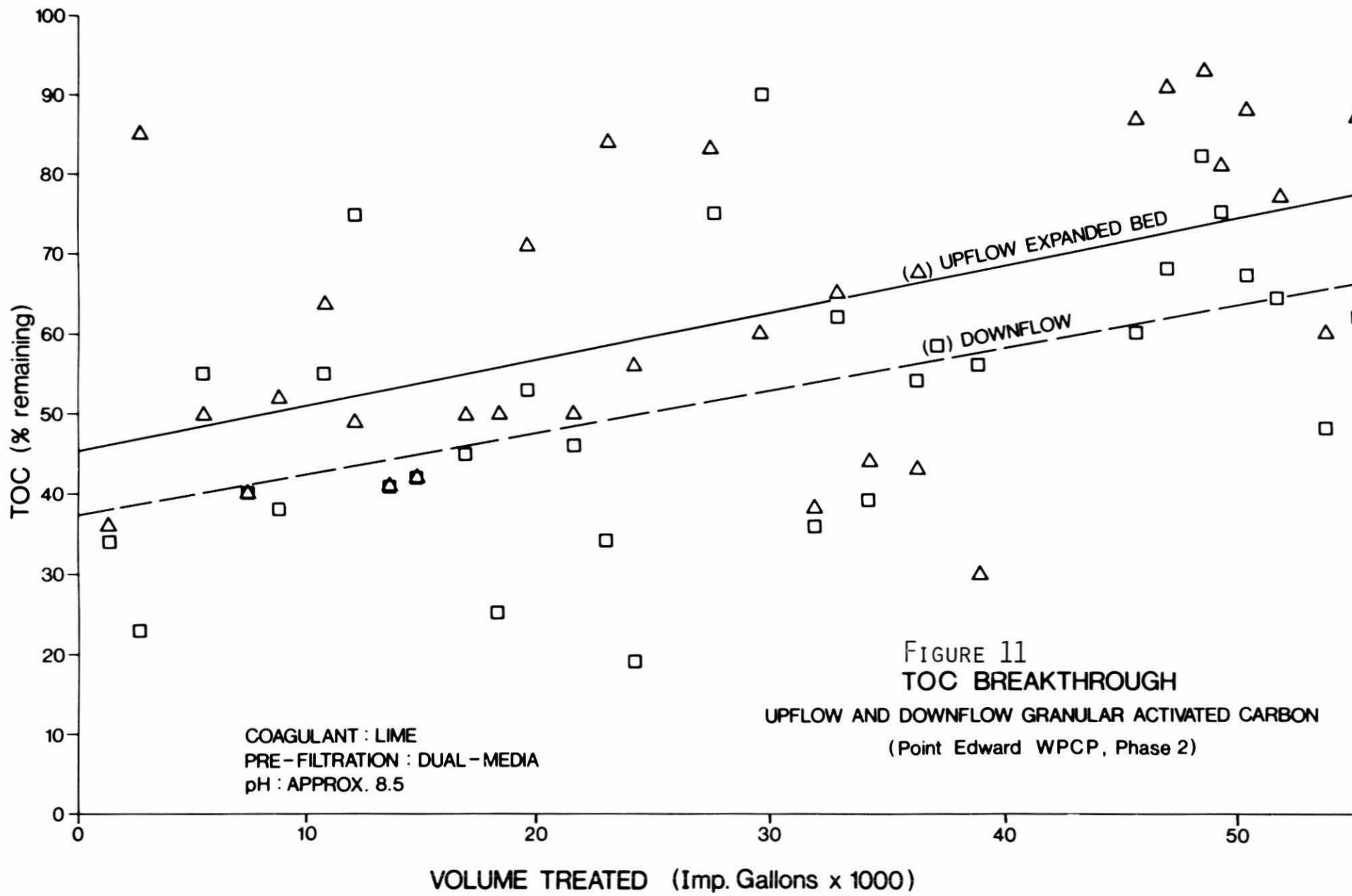
355

FIGURE 11
TOC BREAKTHROUGH
UPFLOW AND DOWNFLOW GRANULAR ACTIVATED CARBON
(Point Edward WPCP, Phase 2)

TABLE 14. GAC IN PHYSICAL-CHEMICAL TREATMENT (PT. EDWARD WPCP)
PERFORMANCE OF DOWNFLOW COLUMNS (PHASE 1)

	TOTAL COD	TOTAL BOD_5	TOTAL TOC	TOTAL TOD	SS	TOTAL P	TURB. (JTU)	TKN	NH_3 N
RAW SEWAGE	1265	288	-	-	999	24	-	-	-
COLUMN INFLUENT	140	70	48	174	67	3.5	25	27	20
COLUMN EFFLUENT	23	16	15	69	11	0.5	4	22	20
PERCENT REMOVAL (over carbon columns)	84	76	69	61	84	85	84	18	0

. All parameters in mg/l except where noted

. Clarification with 300 mg/l alum and 0.25 mg/l polymer, pH approx. 7.0

. Carbon column influent not filtered

TABLE 15. GAC IN PHYSICAL-CHEMICAL TREATMENT (PT. EDWARD WPCP)
PERFORMANCE OF UPFLOW AND DOWNFLOW COLUMNS (PHASE 2)

	TOTAL TOC	TOTAL TOD	TOTAL COD
DOWNFLOW PACKED BED			
Column Influent	37	160	142
Column Effluent	18	81	46
Percent Removal	51	49	68
UPFLOW EXPANDED BED			
Column Influent	37	160	142
Column Effluent	20	101	52
Percent Removal	46	37	63

. All parameters in mg/l except percent removal

. Clarification with 120 mg/l lime, 20 mg/l ferric chloride, pH approx. 8.5

. Carbon column influent filtered by dual-media

adsorption capacities is related to the pH of the wastewater applied,
the higher pH of approximately 8.5 resulting from lime clarification
in Phase 2 being responsible for a reduction in adsorption capacity.

GAC in Physical-Chemical Treatment (Wastewater Technology Centre)

Environment Canada has recently completed the first phase of
a pilot scale study of independent physical-chemical treatment of
domestic wastewater using the high lime process. The pilot plant,
located at the Wastewater Technology Centre, includes chemical coagulation
and clarification at pH 11.5, two-stage recarbonation, pressure sand
filtration, and activated carbon adsorption. The carbon adsorption
units are thirty inch diameter pressure vessels, nine feet in height.
Two columns were operated in series in downflow mode, the first con-
taining 272 kg (600 pounds) of 8 x 30 mesh size carbon and the second
328 kg (720 pounds) of 12 x 40 size. The carbon columns were operated
over a period of approximately three and one half months at various
hydraulic loads. The total quantity of wastewater treated was 4,080
bed volumes [4725 m^3 (1,250,000 gallons)].

Phase 1 of the study was intended primarily to monitor the
coagulation, clarification, recarbonation, and filtration processes.
Carbon adsorption was added to this stage of the study to obtain
preliminary data and to identify operational problems. The hydraulic
load applied to the columns was varied to accommodate studies being
undertaken on upstream units. No activated carbon was replaced during
these experiments. Consequently, the results obtained from carbon
column monitoring in this preliminary study do not represent normal
operating conditions.

The carbon columns were lightly loaded with wastewater
constituents as a result of the removal efficiency of upstream units.
Over five experimental runs the average influent strength was 10 mg/l
total BOD_5, 53 mg/l total COD, 16 mg/l soluble TOC and 9 mg/l suspended
solids. The hydraulic loads imposed on the carbon beds ranged from
0.0007 m^3/m^2h (1.0 gpm/ft^2) to 0.003 m^3/m^2h (4.1 gpm/ft^2) under
steady-state conditions and peaked at 0.004 m^3/m^2h (5.3 gpm/ft^2) with

a diurnal loading variation. Over the duration of the five experiments the carbon columns produced the following average effluent concentrations and percent removals: 8 mg/l (20%) total BOD_5, 31 mg/l (42%) total COD, 10 mg/l (38%) soluble TOC, 6 mg/l (33%) suspended solids. Tables 16, 17, 18 and 19 list the mean influent, effluent and percent removals for these and other parameters for each operating condition. Figure 12 illustrates the change in soluble TOC removal over the five experiments.

Grab samples taken between the two columns were used to estimate the performance of the individual units. The lead column was found to have been effectively exhausted early in the experimental programme and totally exhausted before the completion of the programme. The effective capacity of the lead column, determined by a total BOD_5 effluent concentration breakthrough in excess of 15 mg/l, occurred at a total throughput of 1,000 bed volumes [536 m^3 (142,000 gallons)] and corresponded to an adsorption capacity of 0.025 gram of soluble TOC per gram of carbon. The total capacity of the first column, as determined by the first appearance of consistently zero or negative removals, occurred after it had treated 5,400 bed volumes [2900 m^3 (767,000 gallons)]; the soluble TOC capacity at this loading was 0.096 gram per gram of carbon. The estimate of total or ultimate adsorptive capacity may have been influenced by the 0.004 m^3/m^2h (4.1 gpm/ft^2) loading rate which was applied at 2835 m^3 (750,000 gallons) of throughput. At the end of the experiment, after 4725 m^3 (1.25 million gallons) of wastewater had been treated, the total amount of soluble TOC removed by both carbon beds was 0.068 gram per gram of carbon (4,080 bed volumes, both columns considered). In comparison to the adsorptive capacity of the lead column, this observation suggests that the total adsorption system had exceeded its effective capacity and was approaching its ultimate capacity when the experiment was terminated. The effective capacity cannot be verified because the influent total BOD_5 did not exceed 15 mg/l during the latter part of the study. However, the system may have been near total exhaustion as indicated by the reduced removal efficiencies and increased effluent concentrations of total and soluble BOD_5 and soluble TOC in the last experimental run.

TABLE 16. GAC IN PHYSICAL-CHEMICAL TREATMENT (WTC)
AVERAGE PROCESS PERFORMANCE - BOD_5

PHASE	FLOW (Igpm)	LOAD (gpm/ft^2)	SUSPENDED			DISSOLVED			TOTAL		
			INF.	EFF.	% REM.	INF.	EFF.	% REM.	INF.	EFF.	% REM.
1A	10	2.0	-	-	-	-	-	-	14	10	29
1B	13	2.6	1	3	-200	9	5	44	10	8	20
1C	5	1.0	3	1	67	7	5	29	10	6	40
1D	20	4.1	1	1	0	7	7	0	8	8	0
1E	*		1	1	0	8	7	12	9	8	11
1A to 1E			2	1	50	8	7	12	10	8	20

TABLE 17. GAC IN PHYSICAL-CHEMICAL TREATMENT (WTC)
AVERAGE PROCESS PERFORMANCE - COD

PHASE	FLOW (Igpm)	LOAD (gpm/ft^2)	SUSPENDED			DISSOLVED			TOTAL		
			INF.	EFF.	% REM.	INF.	EFF.	% REM.	INF.	EFF.	% REM.
1A	10	2.0	12	7	42	53	25	53	65	32	51
1B	13	2.6	25	9	64	42	24	43	67	33	51
1C	5	1.0	4	4	0	44	24	45	48	28	42
1D	20	4.1	8	2	75	44	34	23	52	36	31
1E	*		6	2	67	39	29	26	45	31	31
1A to 1E			9	4	56	44	27	39	53	31	42

* Diurnal load variation: Max: 26 Igpm, 5.3 Igpm/ft^2; Min: 2.2 Igpm, 0.4 Igpm/ft^2; Avg: 9.8 Igpm, 2.0 Igpm/ft^2

TABLE 18. GAC IN PHYSICAL-CHEMICAL TREATMENT (WTC)
AVERAGE PROCESS PERFORMANCE - DISSOLVED TOC AND SUSPENDED SOLIDS

PHASE	FLOW (Igpm)	LOAD (gpm/ft²)	DISSOLVED TOC			SUSPENDED SOLIDS		
			INF.	EFF.	% REM.	INF.	EFF.	% REM.
1A	10	2.0	17	8	53	14	10	29
1B	13	2.6	18	9	50	9	7	22
1C	5	1.0	12	7	42	12	5	58
1D	20	4.1	13	11	15	7	6	14
1E	*		17	13	24	6	5	17
1A to 1E			16	10	38	9	6	33

TABLE 19. GAC IN PHYSICAL-CHEMICAL TREATMENT (WTC)
AVERAGE PROCESS PERFORMANCE - PHOSPHORUS, HARDNESS, ALKALINITY

PHASE	FLOW (Igpm)	LOAD (gpm/ft²)	TOTAL PHOSPHORUS			TOTAL HARDNESS			TOTAL ALKALINITY		
			INF.	EFF.	% REM.	INF.	EFF.	% REM.	INF.	EFF.	% REM.
1A	10	2.0	0.68	0.53	22	259	260	0	124	118	5
1B	13	2.6	0.21	0.10	52	178	172	3	138	122	12
1C	5	1.0	0.28	0.31	-11	137	134	2	137	134	2
1D	20	4.1	0.15	0.19	-27	135	148	-10	94	102	-9
1E	*		0.37	0.28	24	155	150	3	117	114	3
1A to 1E			0.38	0.32	16	175	173	1	112	110	2

* Diurnal load variation: Max: 26 Igpm, 5.3 Igpm/ft²; Min: 2.2 Igpm, 0.4 Igpm/ft²; Avg: 9.8 Igpm, 2.0 Igpm/ft²

FIGURE 12
SOLUBLE TOC BREAKTHROUGH
DOWNFLOW GRANULAR
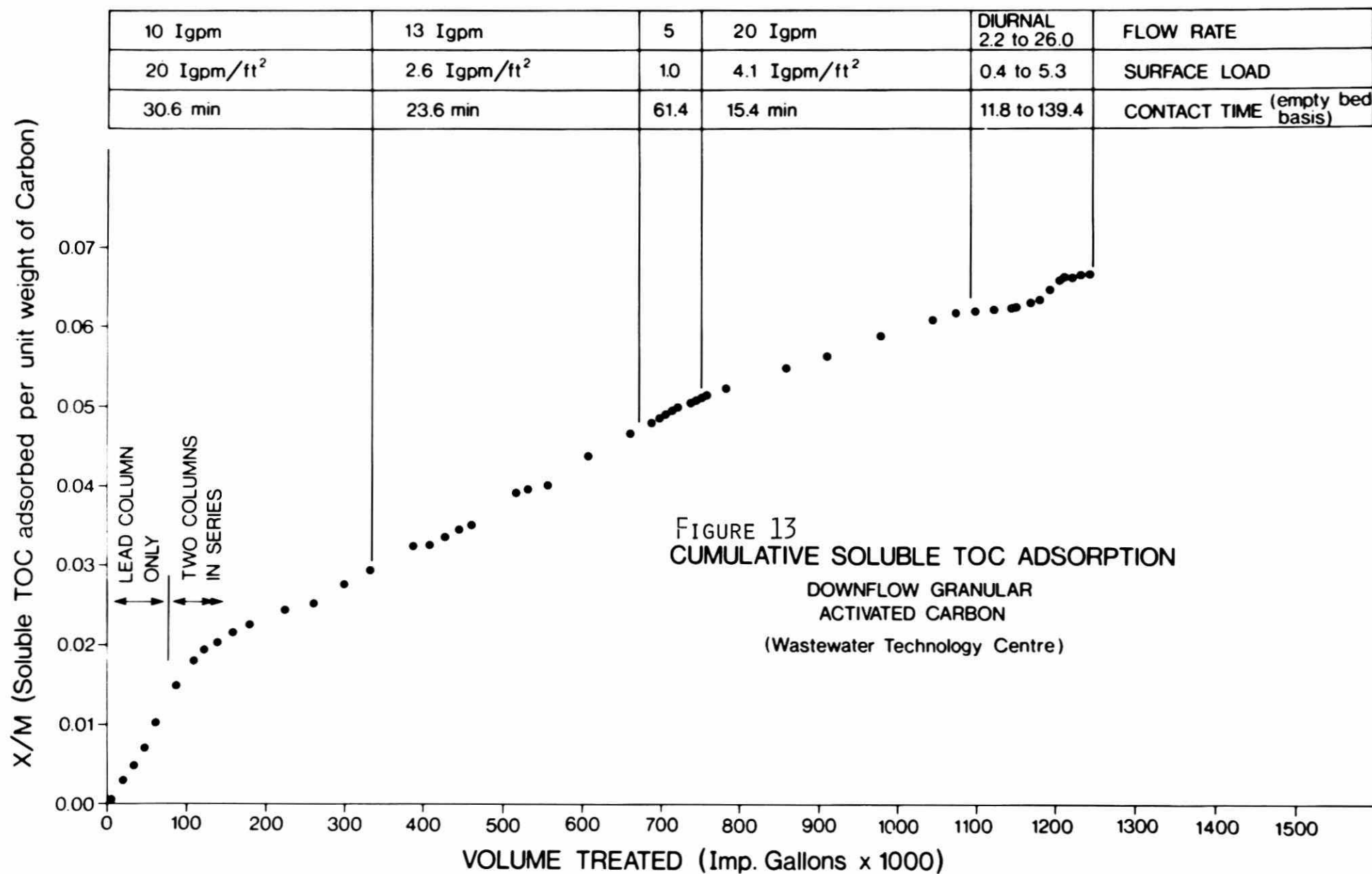ACTIVATED CARBON
(Wastewater Technology Centre)

COAGULANT : LIME
PRE-FILTRATION : SAND
pH : APPROX. 7.0
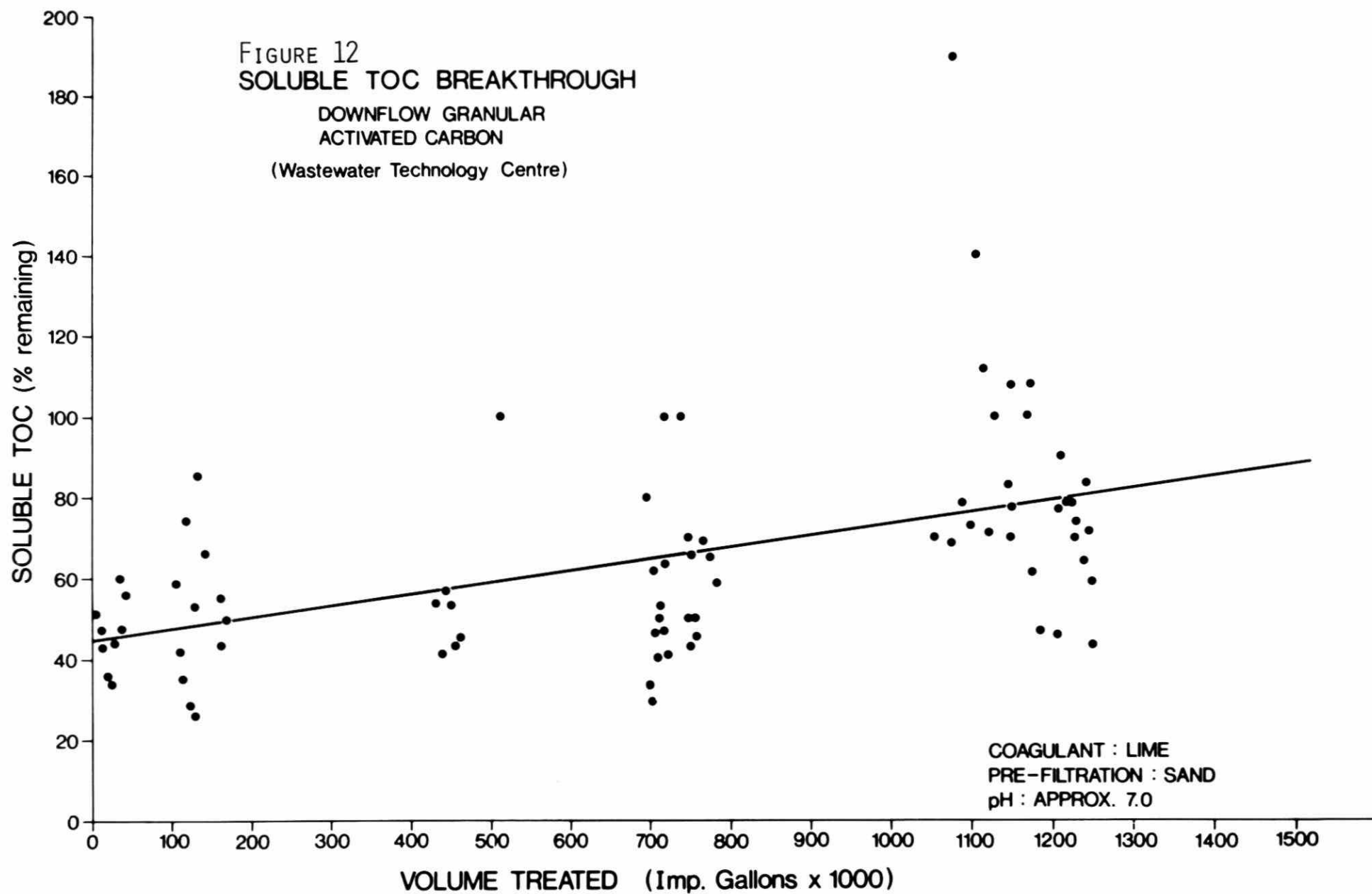
Backwashing of the carbon columns was performed manually at a pressure loss of approximately 104 kN/m^2 (15 psi). The duration of backwash was approximately 15 minutes at sufficient flow to produce a 6% to 8% bed expansion (expansion estimated from carbon manufacturer's data). Higher applied backwash rates resulted in rapid and severe reductions in the actual flow rate possibly as a result of plugging of the upper distribution header screens. This problem has been tentatively attributed to the fact that carbon fines were not manually removed from the bed before operation of the units. It is also possible that the carbon beds pistoned to the top of the columns when backwash was initiated. The frequency of backwashing was determined by the rate of pressure loss increase which is a function of the hydraulic load and solids load imposed upon the carbon beds. At hydraulic loads of 0.0015 and 0.002 m^3/m^2h (2.0 and 2.6 gpm/ft^2) approximately 185 m^3 (49,000 gallons) were treated per cycle. At 0.0015 m^3/m^2h (2.0 gpm/ft^2) backwashing of the lead column was required every three and one half days on the average; at 0.002 m^3/m^2h (2.6 gpm/ft^2) the frequency was two and one half days. At the lowest sustained hydraulic load of 0.00075 m^3/m^2h (1.0 gpm/ft^2) approximately 360 m^3 (95,000 gallons) were treated per cycle with a backwash frequency of 13 days. At the highest loading of 0.003 m^3/m^2h (4.1 gpm/ft^2) the quantity of water treated per cycle and the backwash frequency were reduced to 128 m^3 (34,000 gallons) and one day. The second carbon column was generally backwashed whenever the lead column was backwashed although this was not always necessary particularly at the lower hydraulic loads.

Hydrogen sulphide production in the carbon columns was observed after approximately 10 days of operation or 529 m^3 (140,000 gallons) of throughput. In an effort to control the hydrogen sulphide production it was decided to promote denitrification in the columns by the addition of nitrate. Sodium nitrate was added to the column influent to produce a total NO_3-N concentration of approximately 1.3 mg/l. Within five days the sulphide gas had been eliminated. Biological denitrification under anaerobic conditions resulted in an effluent nitrate concentration of approximately 0.3 mg/l. No attempt was made

to reduce the nitrate dose to produce an effluent free of nitrate. The ammonia and nitrite contents of the wastewater were essentially unchanged through the carbon columns; the average effluent ammonia concentration (as NH_3-N) was approximately 14 mg/l.

The removal of suspended solids by the downflow carbon adsorption units was influenced by the applied hydraulic load as would be expected. A 58% removal was achieved at a surface loading rate of 0.00075 m^3/m^2h (1.0 gpm/ft^2) and at the maximum sustained load of 0.003 m^3/m^2h (4.1 gpm/ft^2) the removal efficiency was 14%. Total phosphorus removal was low (average 16%) and subject to considerable variation. The activated carbon beds did not remove any significant quantity of hardness or alkalinity throughout the five experiments; however, the data suggests that some scale formed on, or was trapped by the carbon beds at low loading rates and was subsequently washed out at 0.003 m^3/m^2h (4.1 gpm/ft^2).

The cumulative soluble TOC adsorption over both carbon columns is plotted in Figure 13. It is interesting to observe the various adsorption rates indicated by the slopes of this graph. The initial slope of the line was created because initially only the lead column was used. When the second column was added the increased M value caused the apparent adsorption rate to decrease. The rate remained relatively constant over most of the first, second and third experimental runs at 1.1×10^{-8} grams soluble TOC per gram carbon per litre (5×10^{-5} lb/lb/1000 gal); it was during this time that the effluent parameter concentrations remained relatively constant. At the 0.003 m^3/m^2h (4.1 gpm/ft^2) loading rate the adsorption rate decreased as did the removal efficiency. During the diurnal loading experiment two observations are significant. There was a sharp increase in the amount of TOC adsorbed at approximately 4536 m^3 (1,200,000 gallons); this increase in adsorption was caused by a shock organic load. The influent soluble TOC, which normally varied from approximately 10 to 20 mg/l, increased to approximately 40 mg/l for a period of about one day, while the effluent concentration increased from approximately 12 to 18 mg/l. Secondly, the removal rate over this experimental run, discounting the shock load, was lower than that experienced at

10 Igpm	13 Igpm	5	20 Igpm	DIURNAL 2.2 to 26.0	FLOW RATE
20 Igpm/ft^2	2.6 Igpm/ft^2	1.0	4.1 Igpm/ft^2	0.4 to 5.3	SURFACE LOAD
30.6 min	23.6 min	61.4	15.4 min	11.8 to 139.4	CONTACT TIME (empty bed basis)

FIGURE 13
CUMULATIVE SOLUBLE TOC ADSORPTION

DOWNFLOW GRANULAR
ACTIVATED CARBON

(Wastewater Technology Centre)

the previous uniform loading rate. The significance of this second observation could be related to the diurnal loading or to the possible exhaustion of the carbon.

GAC in Physical-Chemical Treatment (Bolton WPCP)

The Ontario Ministry of the Environment is presently carrying out full scale studies of granular activated carbon adsorption applied to the effluent of a chemical clarification process at the Bolton Water Pollution Control Plant. Raw sewage to which alum and possibly a polyelectrolyte have been added as coagulants will receive primary clarification and then be directed to the carbon contactor system. This system consists of two five foot diameter columns in series each containing 1818 kg (2400 pounds) of carbon. Product water from the system is stored in a 19 m^3 (5000 gallon) tank and is used to automatically backwash a column when the pressure drop across that column exceeds a preset value. The system has been operated for brief periods to study operational conditions and identify potential problems. During these periods, primary effluent without chemical conditioning was used as feed for the system. Results of grab samples taken with the columns operating at the design flow rate during start-up operations, show a decline in treatment efficiency even for the small volumes treated. The degree of treatment was inadequate to produce secondary effluent quality in terms of the soluble organic parameters. Although some improvement in primary effluent quality is anticipated with the addition of chemical coagulants to the raw sewage, the preliminary results indicate that a reduction of the hydraulic rate or an increase in bed depth may be required to achieve secondary effluent quality.

Discussion - GAC

The effective adsorption capacity of a granular activated carbon bed, under normal operating conditions, was not determined in any of the experiments described above. The adsorption units applied to a secondary effluent in the Bolton experiment did not exhibit a

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breakthrough of unacceptable pollutant concentrations during the treatment of 11,000 bed volumes. The carbon columns treating a filtered high pH primary effluent in the Newmarket experiment could not produce an acceptable effluent quality, but did not produce a definite effluent breakthrough pattern during the treatment of 5,300 bed volumes of wastewater. In the first phase of the Point Edward experiment, 3,600 bed volumes of unfiltered neutral pH primary effluent were treated without any indication of breakthrough. The effluent quality produced in this experiment was relatively good consisting of average concentrations of 16 mg/l BOD_5 and 11 mg/l suspended solids. In the second phase of the Point Edward experiment the feed to the carbon columns was filtered but the pH was raised to approximately 8.5; the effluent quality deteriorated relative to the first phase, but no definite breakthrough was observed during the treatment of 4,600 bed volumes. The carbon columns at the Wastewater Technology Centre were operating on a relatively high quality influent under varying hydraulic loads and no effluent breakthrough was identified.

The Wastewater Technology Centre experiments with downflow granular activated carbon have demonstrated that shock organic loads increase the adsorptive capacity of the carbon by exposing it to higher than normal solute concentrations. This response to loading is predicted by the standard adsorption isotherm test. The use of granular carbon as a tertiary process in the Bolton WPCP demonstrated that the extent to which shock organic loads influence the effluent quality is determined by the remaining adsorptive capacity of the carbon. Therefore, as a carbon bed nears exhaustion, shock organic loads are likely to result in unacceptable effluent quality.

The comparison of fixed bed and expanded bed adsorption processes undertaken at the Point Edward WPCP indicated that expanded beds provide slightly less efficient removal of total parameters, when treating a clarified and filtered wastewater. However, the use of upflow beds reduces the complexity of the equipment and the operational expense associated with frequent backwashing of downflow beds.

Surface loading rates of up to 0.003 m^3/m^2h (4 gpm/ft^2) in downflow adsorption units did not produce unacceptable headlosses or

high effluent suspended solids concentrations. Contact times are another significant factor; on an empty bed basis the contact times used for the Newmarket and Point Edward experiments were between 17 and 22 minutes. Contact times in the Wastewater Technology Centre experiment varied widely, but the results suggested that the critical minimum contact time was approximately 15 minutes. Earlier laboratory tests resulted in equilibrium contact times of approximately 40 minutes for the 8 x 30 mesh size carbon and 10 minutes for the 12 x 40 mesh size (based on soluble TOC).

The use of continuous cumulative adsorption plots may prove to be an effective means of monitoring carbon adsorption processes and predicting column exhaustion. Unlike plots of influent and effluent data or removal efficiency, the cumulative X/M curve forms a recognizable pattern. The slope of the curve should be a function of influent concentration, contact time, and the extent of exhaustion. If a suitable parameter is chosen, for example soluble TOC in the Wastewater Technology Centre experiment, the effective exhaustion of the bed may be predicted. Considerable experience will be required to test the value of this monitoring procedure.

COSTS

Studies of the use of powdered activated carbon have indicated that the process versions studied are not feasible, although the use of granular activated carbon appears to be a practicable means of improving effluent quality. The work described in this paper was essentially preliminary and cannot be used to derive reliable cost estimates. However, Monti and Silberman (1974) have provided cost information for various processes; the following costs were extracted from their article, converted to Imperial measure, and recalculated to include 20-year amortization at an 8% interest rate. The total costs of activated sludge and physical-chemical processes, with and without granular activated carbon adsorption, are plotted in Figure 14. The processes assumed in this estimate are as follows:
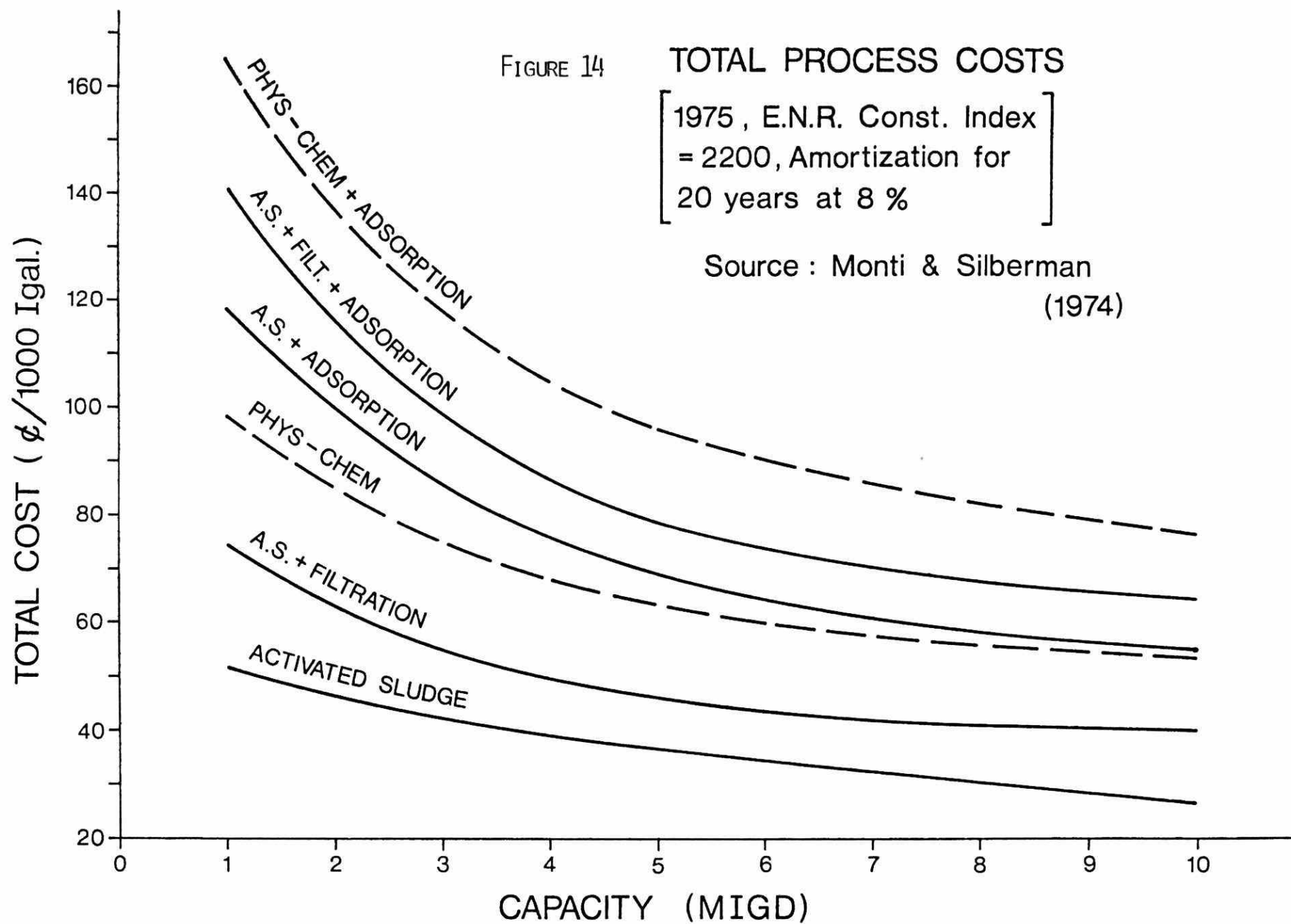
FIGURE 14   TOTAL PROCESS COSTS

$$\left[\begin{array}{l} \text{1975 , E.N.R. Const. Index} \\ \text{= 2200, Amortization for} \\ \text{20 years at 8 \%} \end{array} \right]$$

Source : Monti & Silberman

(1974)

1. Activated sludge estimates include conventional,
   extended aeration, and contact stabilization processes
   with:

   - comminution
   - degritting
   - primary clarification
   - aeration
   - secondary clarification
   - chlorine disinfection
   - sludge disposal

2. Physical-chemical treatment includes:
   - comminution
   - degritting
   - primary clarification
   - chemical coagulation and clarification
   - sand or mixed-media filtration
   - chlorine disinfection
   - sludge disposal

3. Carbon adsorption includes:
   - granular activated carbon
   - contact time of 40 minutes
   - carbon handling systems
   - carbon regeneration
   - carbon attrition of approximately 7% per cycle

4. Nitrogen removal, as ammonia or nitrate, is not included
   in any of the costs shown.

ACKNOWLEDGEMENTS

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REFERENCES

Ahlberg, N.R. and Rupke, J.W.G., "Modes of Activated Carbon Utilization in Municipal Wastewater Treatment", Paper presented at the Technology Transfer Seminar on Physical-Chemical Treatment Activated Carbon Adsorption in Water Pollution Control, Ottawa, 1974.

Beebe, R.L. "Activated Carbon Treatment of Raw Sewage in Solids-Contact Clarifiers", U.S. Environmental Protection Agency, Environmental Protection Technology Series EPA-R2-73-183, 1973.

Burns, D.E. and Shell, G.L., "Physical-Chemical Treatment of a Municipal Wastewater Using Powdered Carbon", U.S. Environmental Protection Agency, Environmental Protection Technology Series, EPA-R2-73-264, 1973.

Hraseova, V., "Carbon Adsorption Studies of Lime Treated Primary Effluent", Ontario Ministry of the Environment, Research Paper No. W2039, Toronto, September, 1973.

Monti, R.P. and Silberman, P.T., "Wastewater System Alternatives: What are they .... and what cost?" Water and Wastes Engineering, Vol. 11, No. 3-6, March - June, 1974.

Wei, N.S. and Le Clair, B.P., "A Pilot Plant Study on the Use of Aluminum Sulphate, Polymer and Powdered Activated Carbon in the Treatment of Raw Municipal Wastewater", Paper presented at the Technology Transfer Seminar on Physical-Chemical Treatment Activated Carbon Adsorption in Water Pollution Control, Ottawa, 1974.